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OPTIMIZING A PETROCHEMICAL WASTE BIO-OXIDATION SYSTEM THROUGH AUTOMATION

Dow Chemical Company

PREPARED FOR:

Robert S. Kerr Environmental Research Laboratory

June 1975

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OPTIMIZING A PETROCHEMICAL WASTE BIO-OXIDATION SYSTEM THROUGH AUTOMATION

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ABSTRACT

Instrumentation and control of an industrial, activated sludge pilot plant was accomplished by development of systems controlling the critical parameters of the process to achieve reliable, high quality effluent. Optimization techniques based on the steady-state and transient models of the activated sludge process were used to determine the minimum volume of the aeration basin required for a specified effluent quality and to predict the transient conditions as a result of step changes in loading.

A pH control system stopped plant operations for the duration of the upset, automatically restoring it when the feed pH returned within operating limits. An automated sampling system, sampling feed and homogenized mixed liquor, monitored the total carbon in both samples. Nutrients, nitrogen and phosphorus (N&P) were added in proportion to the total carbon in the feed, thus maintaining low residual nutrients in the effluent. The sludge recycle flow rate was controlled by a food to microorganisms (F/M) signal, measured as the ratio of total carbon in the feed to that in the mixed liquor. Response time of the F/M control system to a step increase in feed concentration was reduced by 50 to 70 percent, as compared to the uncontrolled system, depending on the amount of excess sludge available for recycle. Chemical flocculants were added in proportion to the turbidity of the biosettler overflow, removing 85 to 98 percent of the suspended solids. Toxic or inhibitory effects of the feed were measured by a Biological Inhibitor Detector, an instrument which measures the oxygen uptake of standard solutions before and after exposure of a bacteria sample to a feed sample and calculates an activity ratio, that had an automated cycle of 60 minutes. The use of the instrument as an upstream sensing device was demonstrated as toxic substances were added to the feed.

The developed on-line control systems are applicable to municipal, industrial, or combined treatment plants. Incorporation of control systems and predictive models in future plants will greatly affect the optimum design and enable more efficient and economic operation.

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16. ABSTRACT

Systems were developed to control the critical parameters of the activated sludge process to achieve reliable, high quality effluent.

An automated sampling system, sampling feed and homogenized mixed liquor, monitored the total carbon in both samples. Nutrients were added in proportion to the total carbon in the feed, thus maintaining low residual nutrients in the effluent. The sludge recycle flow rate was controlled by a food to microorganisms signal, measured as the ratio of total carbon in the feed to that in the mixed liquor.

Toxic or inhibitory effects of the feed were measured by a Biological Inhibitor Detector, an instrument which measures the oxygen uptake of standard solutions before and after exposure of a bacteria sample to a feed sample and calculates an activity ratio. The use of the instrument as an upstream sensing device was demonstrated as toxic substances were added to the feed. The developed on-line control systems are applicable to municipal, industrial or combined treatment plants. A bibliography of 63 references is included.

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SECTION I

CONCLUSIONS

An automated, activated sludge pilot plant was operated on industrial waste to test previously developed control systems in order to optimize the waste control process and to produce a reliable effluent quality. Both fresh water and saline (8 to 10 percent NaCl) waste waters containing glycol were examined as feeds to a completely mixed aeration basin.

The following conclusions were drawn:

1. Automation of an activated sludge system to the degree of sophistication expected of chemical systems control is practical and gives improved efficiency and reliability in operation.

2. Automatic fail-safe operation of the plant was accomplished by the use of individual feedback control loops supervised by an overall shutdown system. The pH control systems, which neutralized the feed within a preset pH control range, operated the feed control valve. Upsets in the feed pH stopped the feed flow, all chemical additions, and sludge wastage for the duration of the upset only, automatically restoring operation when the feed pH returned within operating limits.

3. A sampling system was devised to handle the continuous flow of the feed and take a batch sample from the diluted and homogenized mixed liquor to be injected repetitively into a Total Carbon Analyzer. Operation

of the automated (18-minute cycle) sampling system was satisfactory except for frequent maintenance of the mechanical homogenizer.

4. Nutrient addition proportional to the total carbon in the feed was automated, exhibited a linear response, and performed reliably. The nutrient control system assured microbial viability at all times, and the concentrations of ammonia and phosphorus in the effluent were maintained below 10 ppm and 2 ppm, respectively.

5. On-line food to microorganism ratio (F/M) control is justifiable in activated sludge plants with high yield coefficients where excess sludge is available to increase the bacteria concentration in response to a step increase of organic concentration in the feed.

   The organic loading, or F/M, was measured as the ratio of total carbon in the mixed liquor. The automated (F/M) signal proportionally controlled the fraction of the recycle sludge to be wasted.

   For a fresh water feed, the response time of the F/M control system to a step change in feed concentration was reduced by 65 to 70 percent as compared to the response time of the uncontrolled system having fixed recycle and waste sludge flow rates.

   For a highly saline waste water feed, the response time with F/M control was reduced by about 50 percent, but it was too long (33 hours) due to the low yield of the halophilic bacteria resulting in much less waste sludge available for additional recycle.

6. A Biological Inhibitor Detector was built and operated to monitor the toxic effects of the feed on the activated sludge plant. The automated measurement cycle (60 minutes) compared the oxygen uptake of two standard propylene glycol samples before and after the exposure of the bacteria to a feed sample. Activity ratio as measured by the instrument had a precision of ±5 percent of chart and the oxygen uptake measurement was linear for the range of 200 to 1,000 mg/l of propylene glycol solutions. Tests with toxic substances added to the feed showed that the instrument can detect the presence of the toxin rapidly enough on a repetitive basis to serve as an upstream sensing device in a feed-forward control system.

7. A chemical flocculation control system, using a surface scatter turbidimeter, in a feed-forward mode of proportional control of the alum addition resulted in a high quality effluent. The system was stable, removing 85 to 98 percent of the suspended solids in the biosettler overflow and maintaining the turbidity of the final effluent below 3 Jackson Candle Units (JCU).

8. Neither the Total Carbon Analyzer nor the Total Oxygen Demand Analyzer used to monitor the effluent quality were suitable for operation on highly saline samples. Better accuracy, more precision, and less maintenance was experienced with the Total Carbon Analyzer.

9. Computer programs were developed to calculate plant performance and steady-state and unsteady-state models of the activated sludge process. The steady-state

program was used to determine the minimum volume of the aeration basin required for a specified treatment efficiency at various recycle ratios and recycle sludge concentrations. The unsteady-state program predicted the transient changes in bacteria concentration, aeration basin substrate concentration, and food to microorganisms (F/M) ratio to step changes in loading. The results obtained by this transient model program agreed well with the experimental results.

SECTION II

RECOMMENDATIONS

The developed on-line control systems for the activated sludge process are applicable to sewage, industrial, or combined treatment plants. Based upon the operating experience gained during this investigation, the following recommendations for further development of the control systems and process design are made:

1. The nutrient addition and flocculation control systems could be modified to allow for variations in the feed flow rate by measuring the flow and changing the chemical addition by the multiple of flow times the measured control parameter.

2. The feed-forward inhibition control system is necessary to maintain the viability of the acclimated culture and the high performance of the process. This system is comprised of an upstream biological activity sensing device, an equalization basin, a diversion basin, and pretreatment facilities. The developed Biological Inhibitor Detector should be tested under varying operating conditions as the upstream sensing device. The equalization basin, in addition to dampening the fluctuations in substrate concentration in the influent, provides a time delay together with dilution of a possibly toxic feed during the biological activity analysis period. A diversion basin serves as a holding reservoir of a toxic influent as detected by the Biological Inhibitor Detector and diverted by a control valve or manually as an alarm system is actuated. After determination of the chemical nature of the toxin, it

could be bled back into the process or pretreated by sorption to remove toxic organics, chelation of toxic metal ions, or other indicated process.

3. The food to microorganisms (F/M) control system, to maintain a constant loading of the activated sludge plant, could be modified to include flow measurement. The control system should include an aerated stabilization tank to supply additional sludge recycle in periods of high loading. A pilot-plant study of this concept to determine the best location of the excess sludge tank is recommended. For equal-sized stabilization tanks, the one located in the recycle line would give shorter hydraulic residence time, indicating a higher biota viability while the one located in the sludge waste line would produce a more stabilized sludge, possibly improving the bio-flocculation characteristics.

4. Sampling the influent as it enters the aeration basin allows easy proportional control of the F/M ratio. It is recommended that the sampling of the influent before the equalization basin be studied and that a minicomputer be included and programmed to calcuate the hydraulic effect of the equalization on the feed flow and composition and the F/M control system. This could improve the F/M response time, increasing the sludge recycle rate before the higher loadings affect the aerated biomass.

5. Future plant design procedures will have to incorporate both control systems and predictive mathematical models. These greatly affect the optimum design and enable more efficient and economic operation.

6. There is a need to improve the instruments available to monitor organic concentrations. More reliable operation and reduced maintenance requirements should be the goal, especially when handling highly saline waste water streams.

SECTION III

INTRODUCTION

Aerobic microbial suspensions are used for the removal of a large variety of soluble organic matter from petrochemical waste waters. Such suspensions may consist of dispersed growths as in aerated lagoons or flocculated suspensions of microorganisms, at relatively high concentration, as in the activated sludge process. The activated sludge process is more amenable to control than the aerated lagoon system. The emphasis on control and optimization to produce a reliable effluent quality will continue to make the activated sludge process the principal method of waste water treatment.

For the optimization of the activated sludge process, it is necessary to adjust a number of parameters that are interacting both independently and dependently. The rate-determining steps of the process are: the dissolution of oxygen from the air into the waste liquid, the mass transfer of nutrients from the substrate, and dissolved oxygen to the bacterial cell surface, and the respiration and metabolism of the aerobic bacterial cells. The aggregation or flocculation of the individual microorganisms into matrices to achieve solid-liquid separation could be a controlling parameter for such applications as municipal sewage and food industry waste water.

The metabolic reaction rate has been shown to be controlling for petrochemical waste water treatment (Eckenfelder, 1970). Once suitable conditions of pH, temperature, nutrients, oxygen supply and adequate mixing are supplied, kinetic reactions within the system proceed at rates defined by the nature of the organic matter in the waste water and the activity of the biomass. The controlling parameter of the activated sludge process has always been recognized as the enzymatic activity of the biomass. The success or failure of the activated sludge process is dependent on the control of the food to microorganisms ratio (F/M) and the absence of inhibitors or toxins that retard or destroy the action of the biota.

The controlling parameters of the activated sludge process are well understood. The question is how shall we measure and control these parameters. The common practice of monitoring plant operations with random daily or weekly analysis is inaccurate and clearly inappropriate for control and optimization. Parameters to control the process must be measured on a continuous basis or must be obtained rapidly enough on a repetitive basis to provide updated data for use within the time constant constraints of the process. A real-time response could be achieved when the analytical response time is much less than the unit step detention time.

Valid computer modeling of the activated sludge process exists (Smith and Eilers, 1969). Hardware is available to perform feedback, feed forward, adaptive and optimizing control, in analog or digital execution or a combination thereof.

The problems of automated control of activated sludge plants lie in selecting parameters and developing sensors for these parameters rather than in the analysis and presentation of the accumulated data.

The objective of the reported study was to develop instrumentation to control an industrial activated sludge plant with the same degree of instrument sophistication that controls a continuously operating chemical process. For this purpose, an existing, skid-mounted miniplant was utilized. A simplified flow diagram of the plant is shown in Figure 1. This plant was successfully operated to process a high salt (8 to 10 percent) industrial waste containing polyhydric organics (sucn as glycol and glycerine), extending the activated sludge process into salt concentrations not previously believed possible (Zeitoun, et al. 1971). Chemical flocculation following the bio-settler was necessary to remove the turbidity produced by the halophilic bacteria and was considered part of the process.

Control of the feed pH and manual control of the air supply to the aeration basin were features of the existing plant. The specific control systems developed during the reported study are illustrated schematically in Figure 2 and include:

1. Nutrients Addition Control

   Control of the nutrients addition by sensing the carbon in the feed and adding the nutrients in a ratio to the measured carbon was proposed. The carbon in the feed was to be measured with an organic carbon analyzer. The nutrient solution, containing fixed and known amounts of nitrogen and phosphorus, was to be added in a ratio to insure the bacterial viability and to regulate the amount of residual nitrogen and phosphorus in the effluent, improving both the quality of the effluent and the process economics.

2. Food to Microorganisms (F/M) Ratio Control

   A real-time measure of the loading ratio would allow control of the biomass in the aeration basin and the

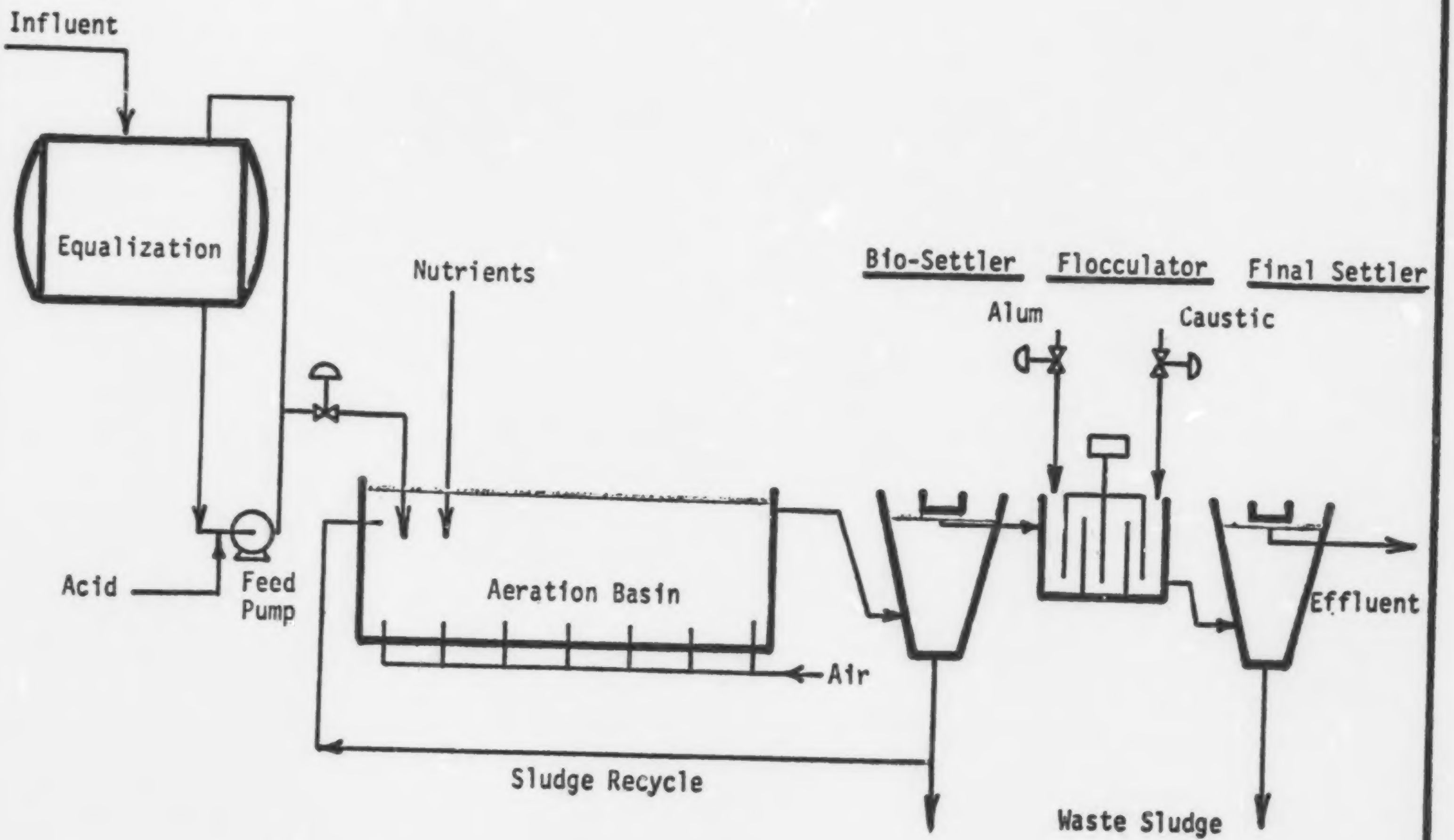


Figure 1. Activated sludge miniplant---flow diagram

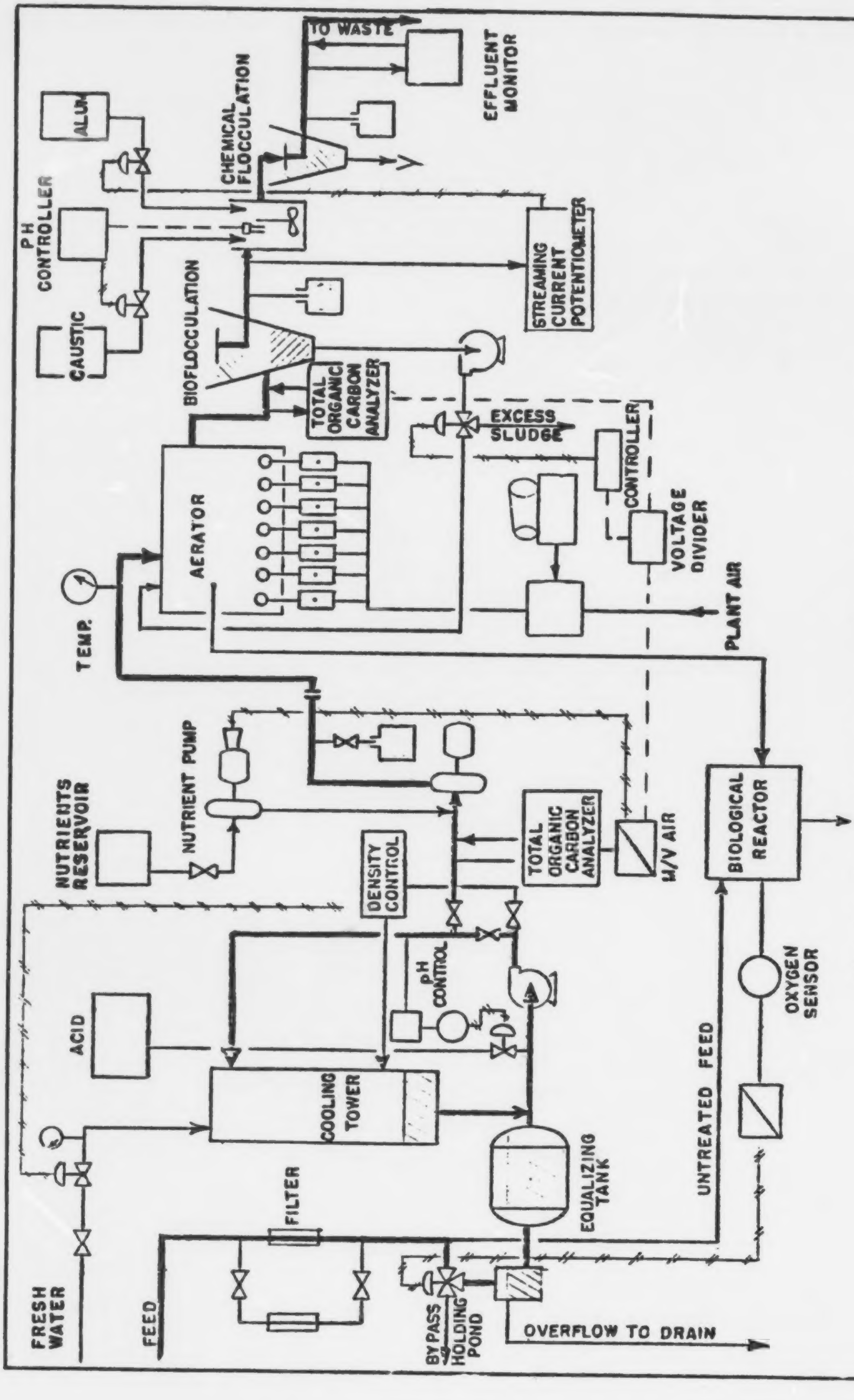


Figure 2. Activated sludge control system

amount of sludge to be wasted and returned to the process. It was proposed to measure the concentration of the bacteria in the overflow from the aeration basin and the concentration of substrate in the feed by a total carbon analyzer. The measured values, as organic or total carbon, would be transmitted to a differential controller. The output of the controller would actuate a control valve on the sludge return line to keep a constant loading ratio.

3. Flocculation Control

A measurement of turbidity in the effluent was proposed to control the amount of alum fed to the flocculator. The pH in the flocculator would be controlled by the addition of sodium hydroxide solution at a level corresponding to the optimum pH for alum flocculation.

4. Biological Inhibitor Detector

An upstream sensing device was proposed to detect toxic loads in the feed in a feed-forward control system. The sensing device, a respirometer or a biological reactor would measure the rate of the biological reaction of the acclimated culture and the feed at regular intervals. When the presence of a toxin was sensed, it would either divert the toxic feed to a holding pond or actuate an alarm system.

It was also proposed to apply a total oxygen demand instrument to monitor the effluent quality and to assess the performance of the plant. The evaluation of the automatic control in increasing the plant efficiency to obtain reliable high quality effluent would be made utilizing a simplified mathematical model of the activated sludge bio-oxidation process.

SECTION IV

ACTIVATED SLUDGE MINIPLANT

The activated sludge miniplant, designed, constructed, and built under Grant No. 12020 EEQ, similar to that described by Mulbarger (1966), was designed for an average flow rate of 0.5 gallons per minute. This miniplant was successfully operated for more than a year on an equalized propylene glycol waste water that contained a high salt concentration (8 to 10 percent NaCl). The results of this testing and a description of the plant are contained in the Final Report submitted in the previous grant (Dow, 1970). The operational and design parameters determined from the miniplant operation were verified in a 150-gallons-per-minute pilot plant utilizing commercial size equipment.

For the present study, the existing plant was renovated, new electrical circuits for the instrumentation were added, and the existing circuits modified for outdoor operation.

PLANT DESCRIPTION

The skid-mounted miniplant was connected to an equalization tank and a cooling tower (if necessary). Details of the unit may be seen in Figures 3 through 9. The major components of the plant, mounted on the skid are:

1. An acid tank, feeding into the suction line of the recycle pump through a pH-controlled automatic valve (Figure 3).

2. A nutrient solution tank supplying nutrients to the aeration basin through a metering pump.

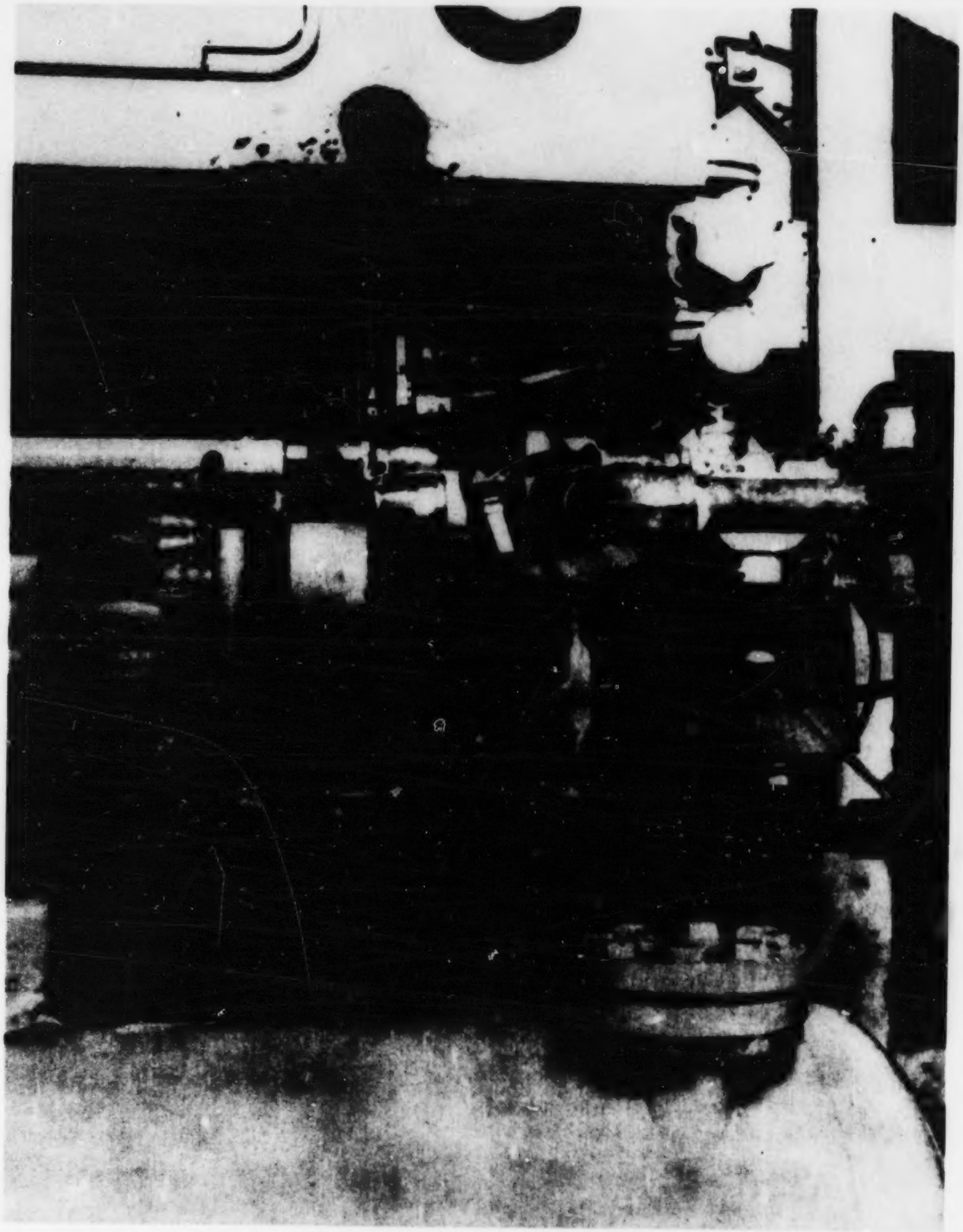


Figure 3. Activated sludge pilot plant--acid and pH control system

Figure 4. Activated sludge pilot plant---total carbon analyzer for nutrients control and sludge recycle control

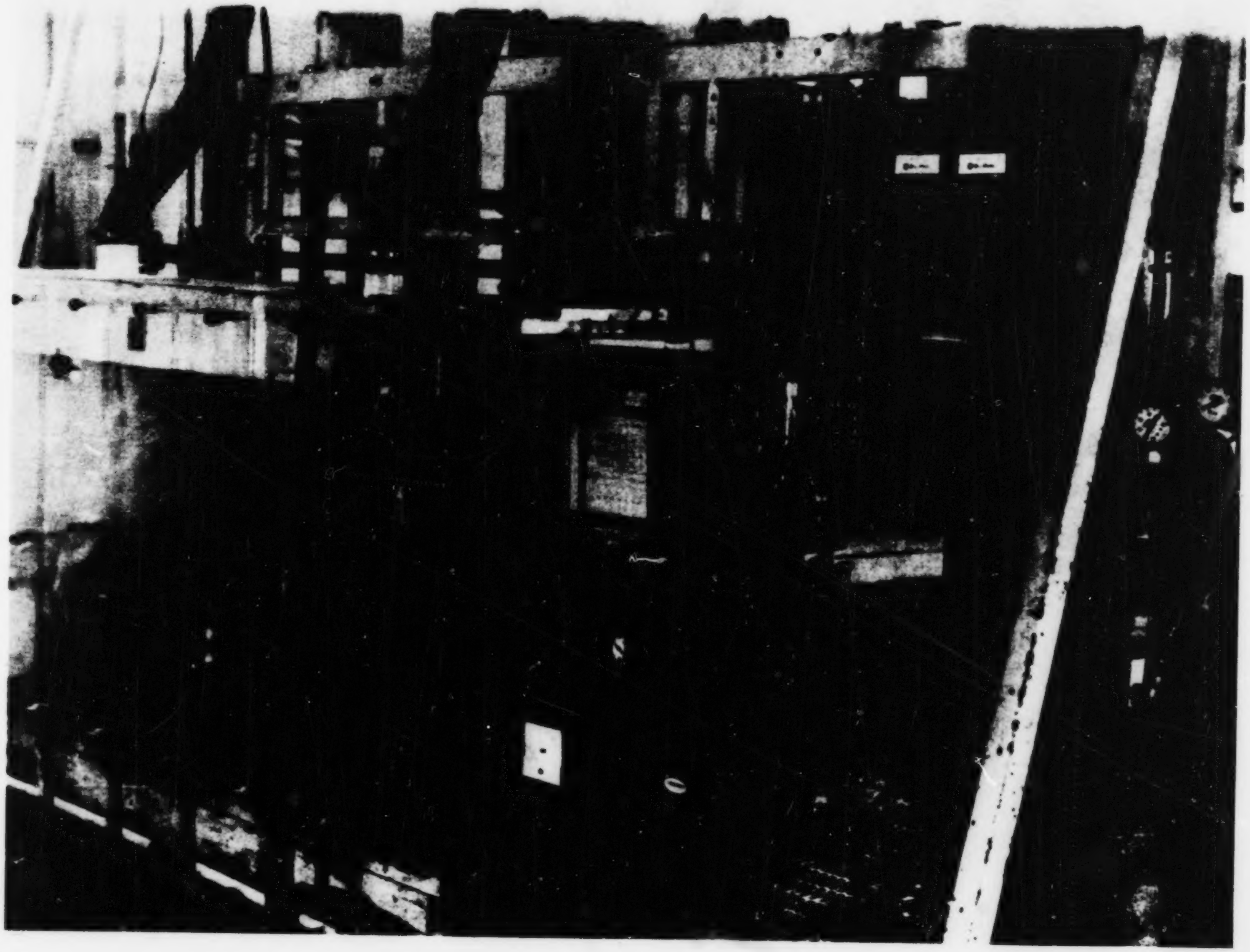


Figure 5. Activated sludge pilot plant---flocculation control and effluent quality monitor

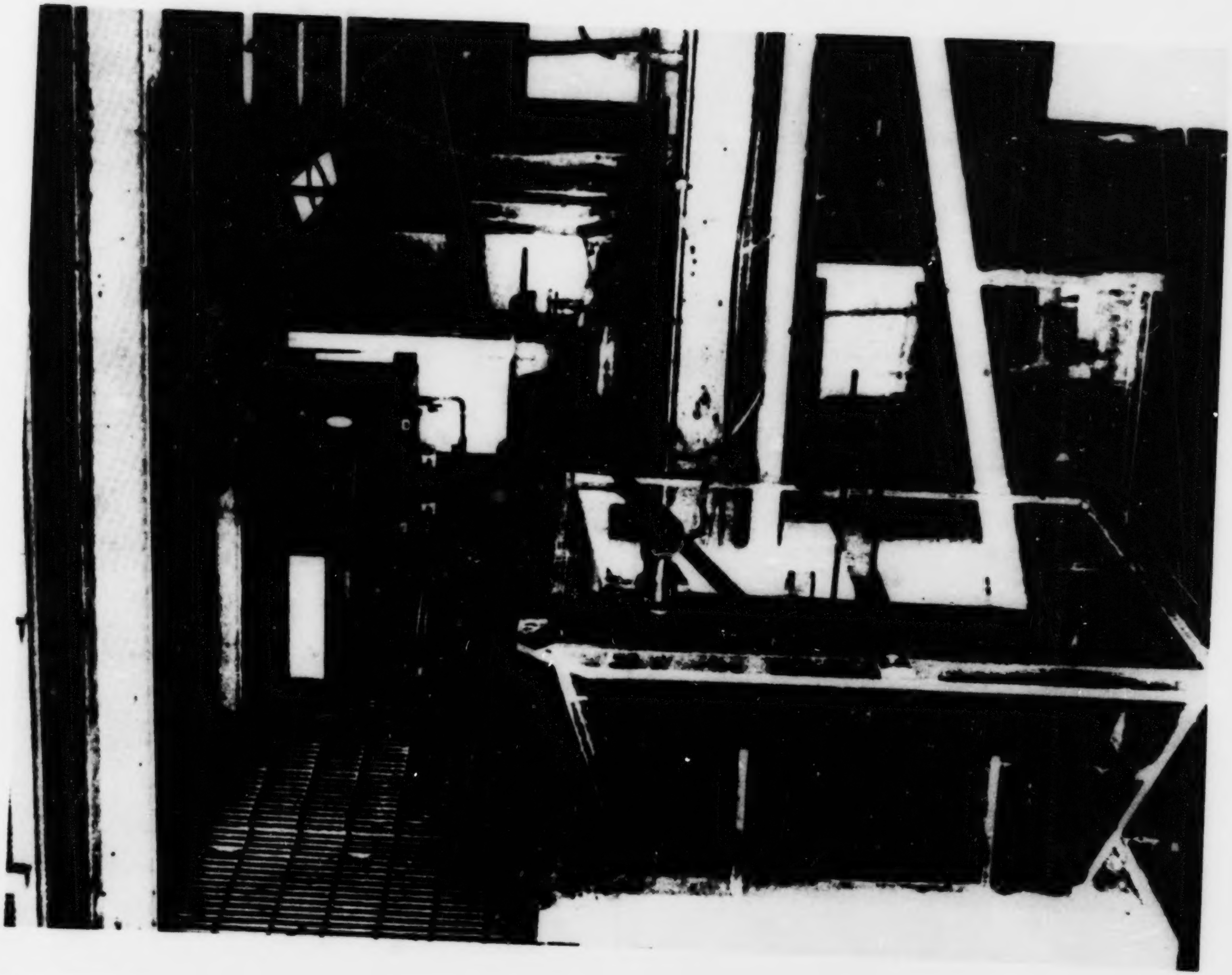


Figure 6. Activated sludge pilot plant---control panel

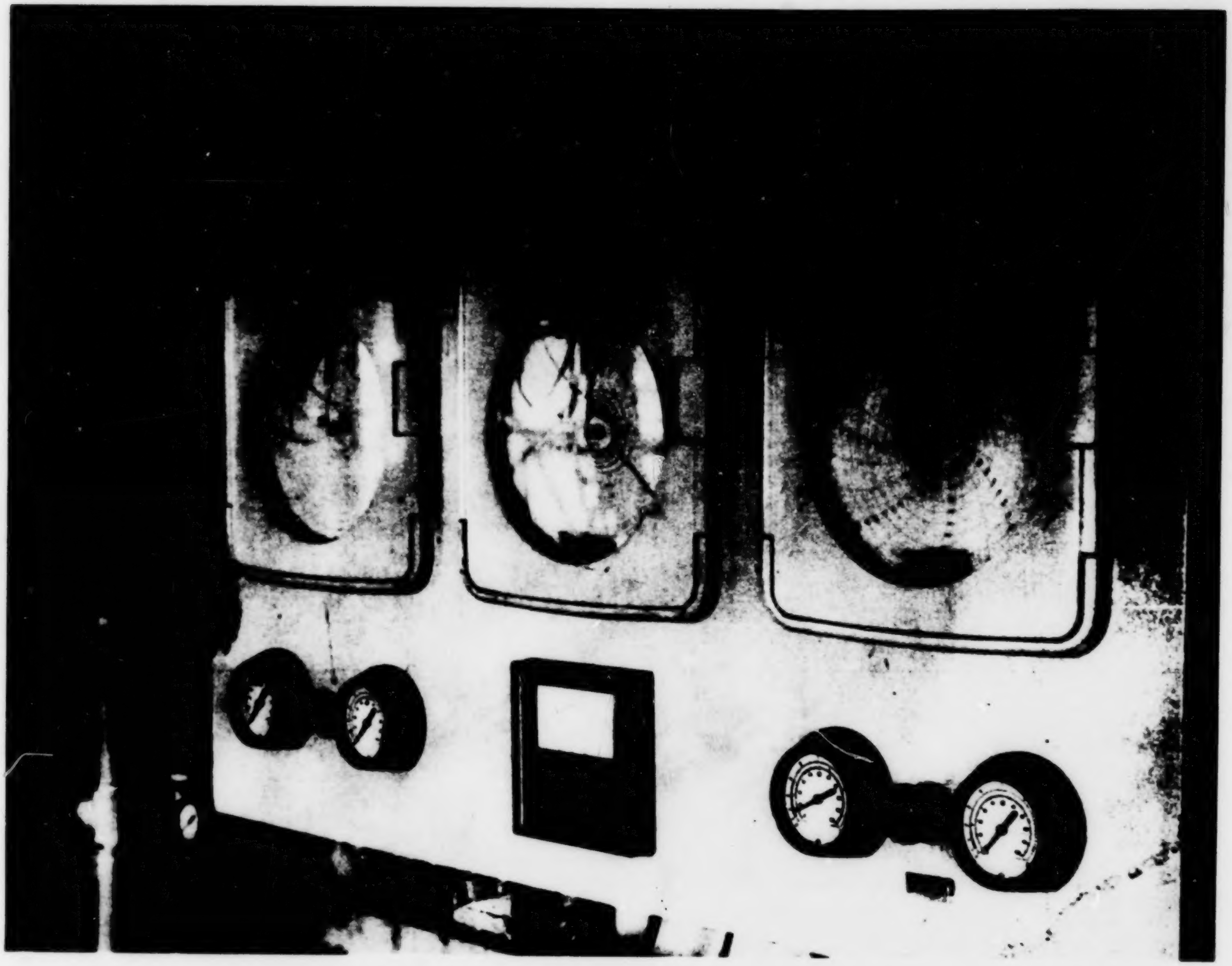


Figure 7. Activated sludge pilot plant---
Plan View

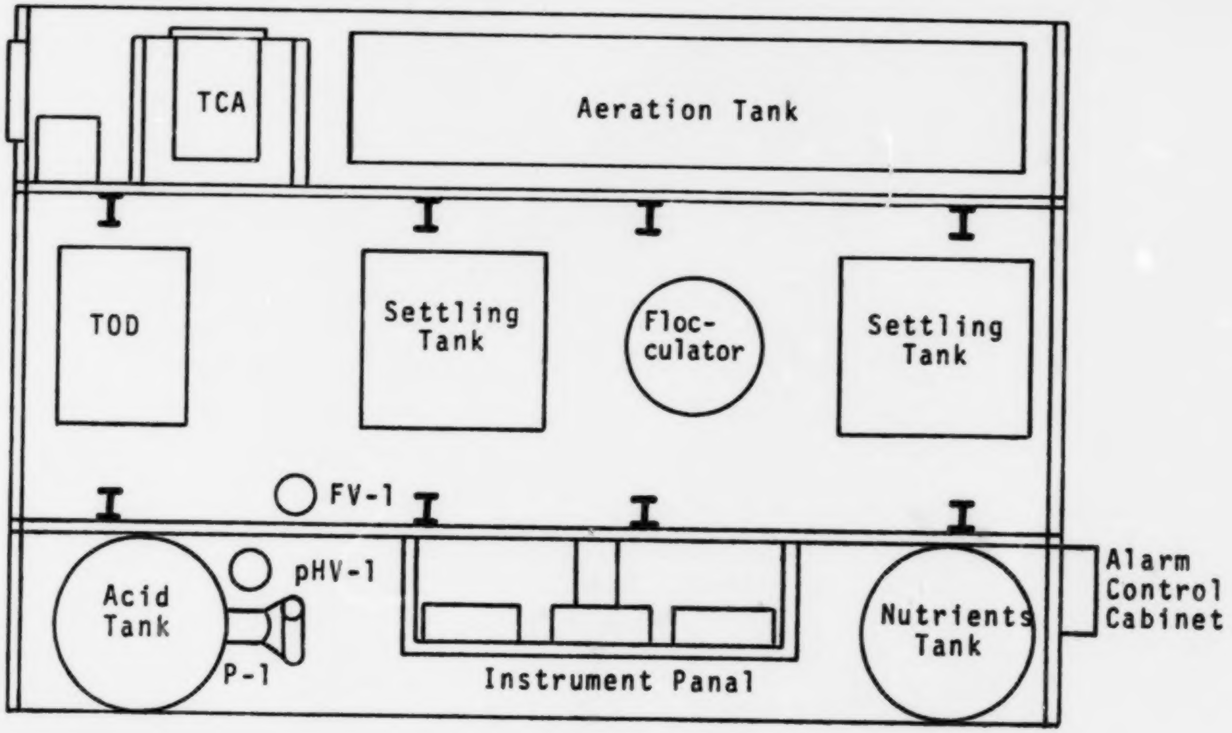


Figure 8. Activated sludge pilot plant---
Side View

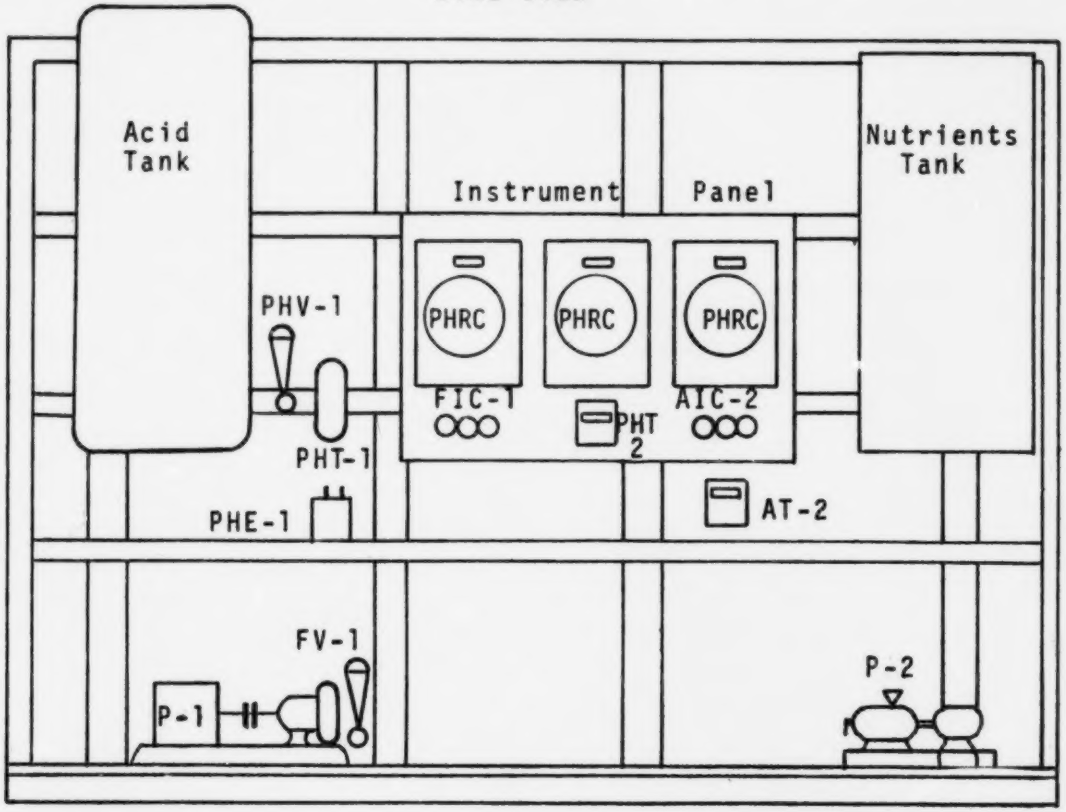
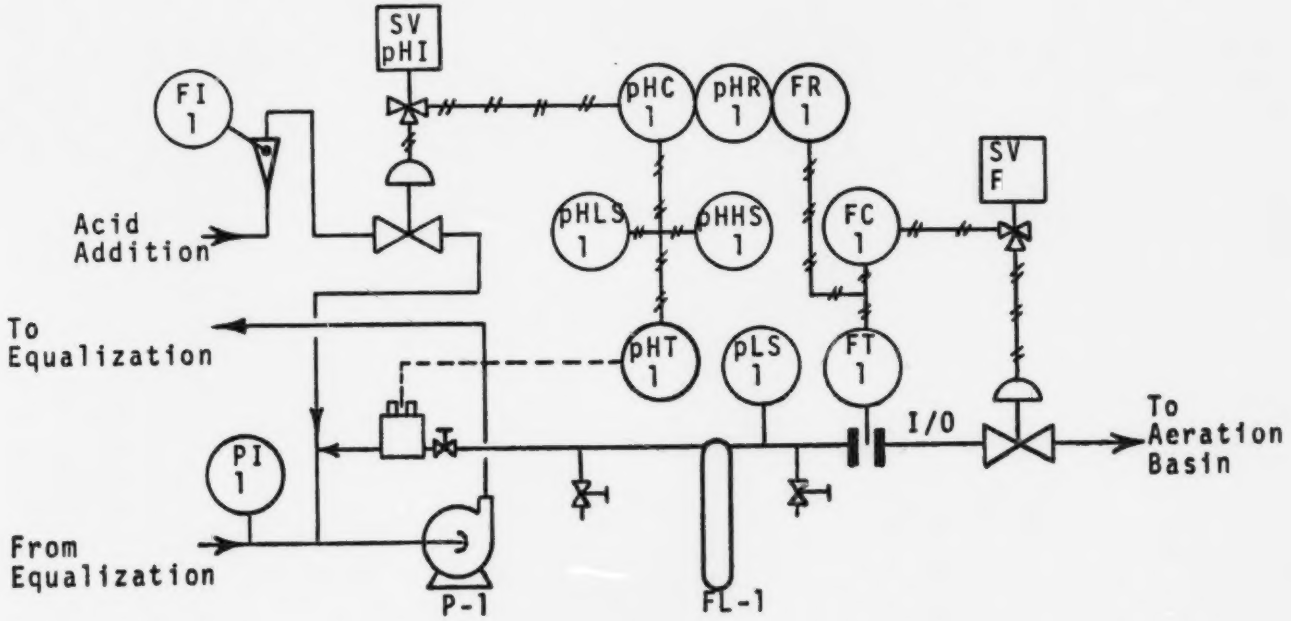


Figure 9. Influent flow and pH controls



3. A 280-gallon aerator tank (Figure 4) made of Plexiglas with six removable partitions was operable as a plug flow reactor or a completely mixed tank. Diffused air was introduced into seven points in the bottom of the tank and controlled by needle valves at the airflow rotameters.

4. Two 50-gallon Plexiglas settlers, each with an adjustable overflow weir and airlift or pump to remove the settled solids from the bottom. The bio-floc from the first settling tank was recycled to the aeration tank and part of the flow wasted as excess sludge. A Total Carbon Analyzer (Figure 4) measured the total carbon of the feed and the mixed liquor from the aeration basin.

5. A flocculator consisting of a 10-gallon glass battery jar with a stirrer fed alum and caustic through control valves and placed between the two settling tanks (Figure 5).

6. A Total Oxygen Demand Analyzer to monitor the total oxygen demand (TOD) of the influent and effluent before and after chemical flocculation. The TOD analyzer was equipped with a five stream selector and sampling filters and pumps.

7. The control panel (Figure 6) included instruments to indicate the major measured parameters of the process. The influent pH and its flow rate were chart recorded. The influent flow rate was set manually at the control board. The sludge recycle was recorded and was manually controllable at the board. The alum flow rate and the turbidity of the final effluent were both

recorded and the flow of the alum solution was controlled at the board. The pH of the final effluent was both recorded and indicated.

A plan view and a side view of the miniplant are shown in Figures 7 and 8, respectively. The control panel instrument legend is described in Table 1.

The influent pH control system was designed for fast response in a well mixed system. By coupling the pH feedback loop around the recycle feed pump, the loop capacity, mixing time and measurement lags were kept to a minimum. The pump also served as the feed pump to the aeration tank and provided the feed at the proper pH independent of the volume and pH of the influent in the equalization tank. The influent flow and pH controls are shown in Figure 9 with the instrument list given in Table 2.

The pH measurement was made with a glass electrode and a solid-state reference electrode fitted in a polypropylene holder of the flow-through type. The electrodes were easily removable for calibration in buffer solutions. The control system was composed of a Bell and Howell solid-state pH-to-air converter, a Foxboro pneumatic recording controller, and an Annin control valve with a special Kynar body and trim and a flow coefficient (C_v) of 0.1 with linear characteristic. The sample flow taken from the pump discharge passed through a ball valve that throttled the sample flow so it was large enough for good mixing at the acid injection point without subjecting the pH electrodes to the excessive pressure of the pump discharge. The velocity distance lag time of the pH control system was about two seconds.

Table 1. CONTROL PANEL, INSTRUMENT LEGEND

Legend	Description
P-1	influent recycle pump
P-2	nutrients proportioning pump
pHRC-1	influent pH control
FIC-1	influent flow control
AIC-1	nutrients flow control
RRC-1	sludge recycle control
pHRC-2	effluent pH control
AIC-2	alum dosage control
AT-2	turbidity transmitter
pHE-1	influent pH electrodes
pHT-1	influent pH transmitter
FV-1	influent flow control valve
FR-1	second pen of pHRC-1
AR-1	second pen of pHRC-2
AR-2	second pen of RRC-1
TOD	Total Oxygen Demand Analyzer
TCA	Total Carbon Analyzer

Table 2. INFLUENT FLOW AND pH CONTROLS, INSTRUMENT LIST

Tag No.	Description	Range	Set Point
FT-1	Foxboro M13AM-MS 2, 0.159 I/O	0-84" w.e.	
FC-1	Moore Products M 55A	0-1 GPM	.5 GPM
FV-1	Annin M1560, 3/4" NPT	C_v = .6	
FR-1	(second pen of pHR-1)	0-1 GPM	
FI-1	Brooks, PVC body		
PI-1	Ashcroft 1279-S	30"-0-30 PSI	
PLS-1	Static-O-Ring #12N-K45	0-16 PSI	4 PSI
pHT-1	Bell & Howell Type 18-155	2-12 pH	
pHC-1	(integral to pHR-1)	2-12 pH	7.7
pHR-1	Foxboro M 40R-A4-2	2-12 pH	
pHV-1	Annin Domotor w/Kynar body and trim	C_v = .1	
pHHS-1	Static-O-Ring #12N-K45	0-16 PSI	pH 8.5
pHLS-1	Static-O-Ring #12N-K45	0-16 PSI	pH 6.5
SV-F, SV-pH-1	Skinner 3-way solenoid valve #V53 LB 2100, 1/4" NPT, 115V coil		

The pH control system had a high and low limit switch that closed the feed valve if the pH got out of control and opened the feed control valve as the control range was regained. The control points were set at any pH limit for automatic control.

The influent flow rate to the aeration basin was controlled with an integral orifice d/p transmitter providing the flow measurement that was recorded on the second pen of the pH recorder.

PLANT OPERATION

Automatic operation of the plant was accomplished with a number of individual feedback control loops supervised by an overall shutdown system. Each individual control loop functioned in its prescribed manner as long as the feed stream pressure and pH were maintained. The electrical schematic of the shutdown system is given in Appendix A.

Operation of the process was initiated by a pushbutton to start the feed pressure pump. A short duration time-delay relay allowed time for pump priming. Once the feed pressure was sufficient, the pH control system automatically added acid to bring the feed pH within the control limits. This was automatically followed by the influent flow control that throttled the feed to the aeration basin at the set value. All forward flows from the aerator and the subsequent equipment were regulated by gravity heads.

The loss of feed pressure caused all chemical additions and sludge wastage to stop until the system was manually reset. An upset in the feed pH stopped the same operations but only for the duration of the upset. Operation was automatically restored when the feed pH returned to operating limits.

A pH bypass switch was provided to inhibit automatic shutdown during periods of attended testing. Two time-delay relays protected this automatic shutdown system from false operation due to electrical transients.

Inidividual feedback control loops were arranged for fail-safe operation. In case of power loss or air supply within a loop, the control element closed.

Details of the start-up procedures, routine maintenance of the instruments and controls systems, and the sampling schedule during operation of the plant are given in Appendix A.

CALCULATIONS OF PLANT PERFORMANCE

A computer program was developed to facilitate routine data handling and to calculate parameters which described the miniplant performance. The output from the program aided evaluation of the proposed control systems and operational decision making. The details for the program are presented in Appendix B.

Data entered into the program represent daily averaged values. Included were Total Oxygen Demand (TOD), Total Carbon (TC), Mixed Liquor Volatile Suspended Solids (MLVSS), turbidity, temperature, pH, and flow data.

The height of waste sludge accumulated in a receiving barrel was entered and converted into an average waste flow rate. A 24-hour average recycle flow rate was calculated from flow data from a one-minute grab sample of both waste and recycle. The sum of the waste and recycle flow was a constant, and it was necessary to calculate this flow since the individual recycle and waste flows were not constant because of the F/M control action.

Among the data tabulated by the program were TOD removal and the formulation of volatile suspended solids. A microorganism balance on the aeration basin allowed calculation of the specific substrate removal and the specific growth rate parameters, which are necessary for the generation of kinetic constants (see Section VI).

Carbon and oxygen material balances were also provided by the program. Insertion of program logic to evaluate oxygen transfer efficiency was optional.

The mean cell residence time, commonly called the sludge age, was a useful operational parameter. It was calculated on a daily basis as the average microbial mass (as MLVSS) in the aeration basin divided by the total mass flow of wasted microorganisms and biota lost through the bio-settler overflow.

The program out put included calculation of the sludge volume index (SVI) based on the measured, 30-minute, mixed liquor settling time in a 1,000-ml graduated cylinder.

The performance of the control systems, presented in terms of daily averaged values, included calculation of carbon-to-nitrogen-to-phosphorus ratios for the nutrients control system, alum usage for the flocculation control system, and loadings for the F/M control system.

SECTION V

FOOD TO MICROORGANISMS (F/M) CONTROL SYSTEM

The most important parameter in optimum control of the activated sludge process is the loading ratio, or the food to microorganisms ratio. Good F/M control will result in a high quality effluent and a sludge with good settling properties.

For a given hydraulic loading, F/M is determined by the ratio of the concentrations of the feed stream and the aerated bio-mass. The method used in the pilot plant was to make on-line total carbon analysis of each of the two streams, convert these to stored values, and adjust the recycle flow to the aeration basin by proportional F/M control.

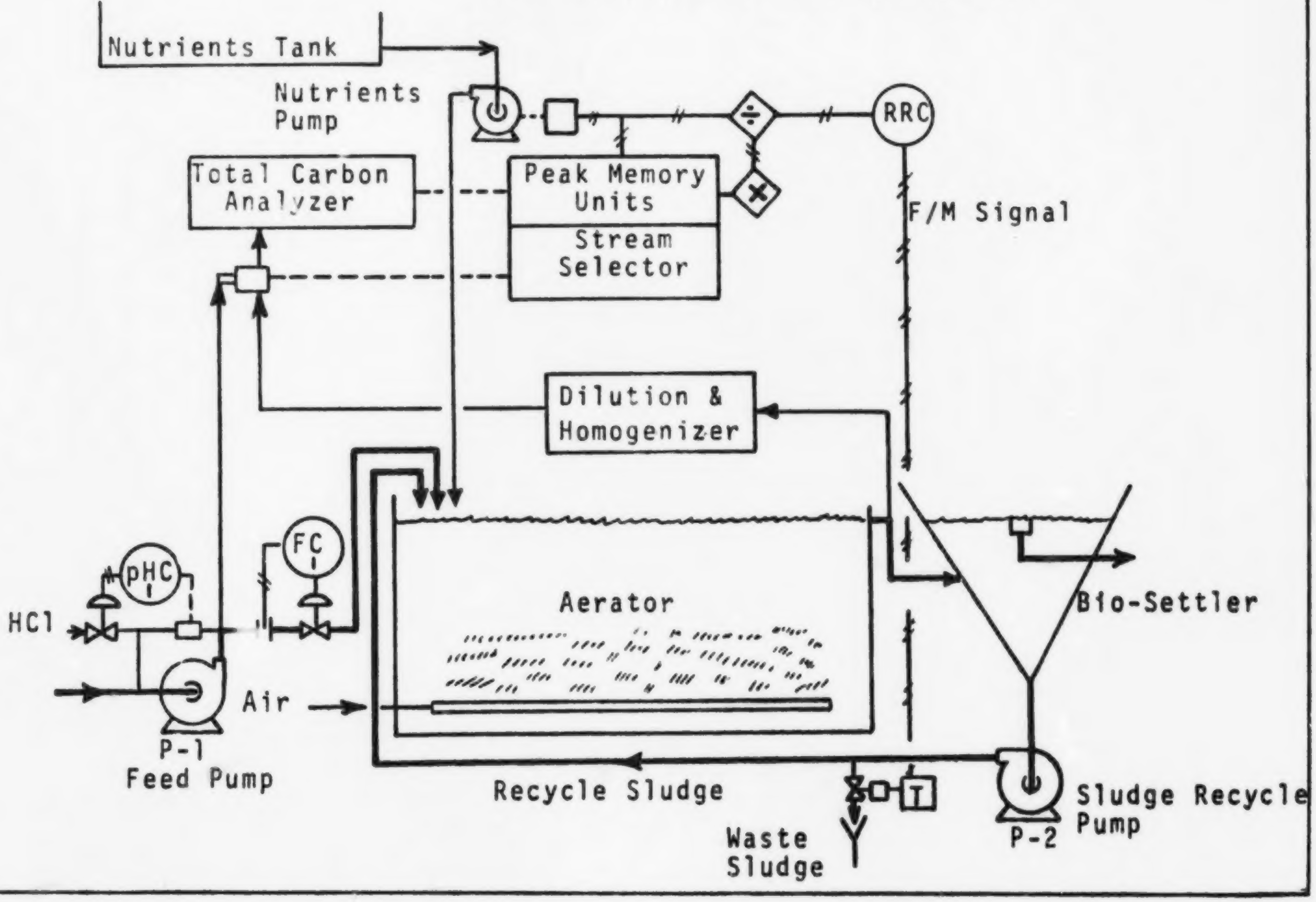
SYSTEM DESIGN

The food to microorganisms control system is shown schematically in Figure 10, together with the nutrients addition control system.

An Ionics Model 1212 Total Carbon Analyzer was used to analyze the feed and the mixed liquor samples. Sample injection was provided through a slide valve with a 40-microliter sample volume alternated with a 100-microliter distilled water wash. A MSA Model 300 IR unit, sensitized for carbon dioxide, was the sensor portion of the analyzer. The recorder housed the analyzer time controls and the combusion tube controls. The automatic zero circuitry was added to the rear of the infra-red unit housing.

The analysis-cycle time was set by the speed of the program cam timer which controlled the operation of the sample

Figure 10. Activated sludge pilot plant---
Nutrients control and F/M control



injection valve and the auto-zero circuitry. Added to this time was the control necessary to synchronize the peak picker, peak transfer, and peak picker reset circuitry with the analysis signal. Also added to the timer was the sample stream transfer controls.

Sample Handling System

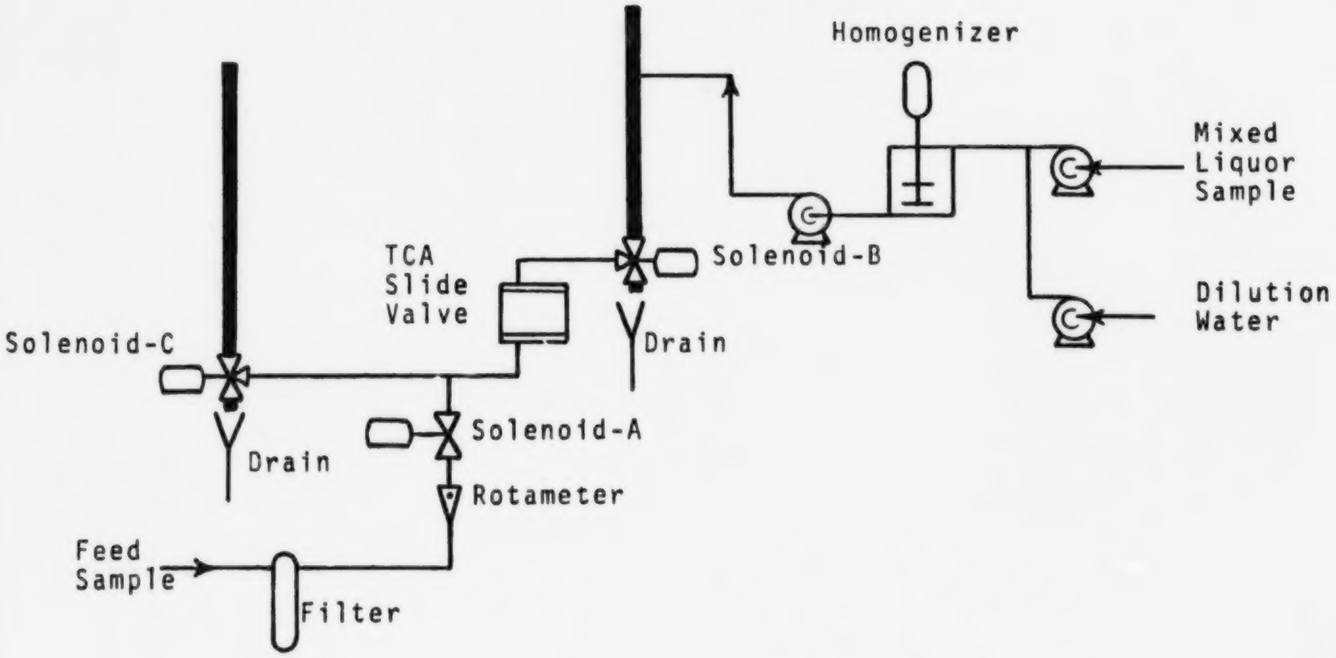
The sample handling system of the Ionics Model 1212 Total Carbon Analyzer was modified to accomodate two sample streams, a continuous flow and a batch sample. The filtration and bypass components of the original analyzer were removed (Figure 11).

A continuous sample of feed passed through a 3-micron Cuno filter, through a valved rotameter, and through a two-way manually-opened solenoid valve (A). It then passed through the Total Carbon Analyzer (TCA) sample injection slide valve to a special three-way solenoid valve (B) and on to drain. The feed sample was blocked from three-way solenoid valve (C).

The mixed liquor typically contained from 1,000 to 5,000 milligrams per liter of suspended bio-solids, whose random floc size was too large to obtain a reproducible aliquot of the 40-microliter size needed for analysis. Hence, homogenization of a larger aliquot was performed in order to obtain a representative sample. A Virtis 45 homogenizer was used, operating at about 10,000 rpm.

A 1:1 dilution of the mixed liquor sample aided in the homogenization step. Two timers operated sequentially to first operate pumps providing the aliquot and next operate the

Figure 11. F/M control sampling system



homogenizer. Upon completion of the homogenization time, another sample pump emptied the homogenizer contents into the stand-pipe above solenoid (B), which allowed deaeration of the sample before analysis. All three solenoid valves were energized at that moment: two-way solenoid valve (A) was closed, and the drain connections of valves (B) and (C) were closed. The homogenized sample passing through the stand-pipe above valve (B) flowed by gravity through the sample injection slide valve and into the stand-pipe above solenoid valve (C) until the stand-pipe levels approached equilibrium. Flow in this direction flushed the prior feed sample from the injection valve.

Upon injection of the homogenized sample, all of the solenoid valves were de-energized, restoring the flow of the continuous feed sample stream through the system. Meanwhile, the batch sample in both stand-pipes was drained.

Energizing the solenoid valves from a power source outside the automatic stream-swtiching controls permitted the addition of a standard solution to the stand-pipes for calibration of the instrument.

Peak Picker and Signal Memory

The signal to the recorder of the Total Carbon Analyzer was 0 to 5 millivolts. By adding a retransmitting slidewire inside the recorder, a signal range of 0 to 5 volts was obtained. The simplest form of peak picker consisted of a single diode and capacitor. The charge on the capacitor was proportional to the height of the peak signal applied. Quick transfer of this charge to a sample and hold amplifier provided storage of that signal value for extensive periods of time. Following transfer of the charge, a relay was energized to discharge the capacitor, preparing it for the next peak to be applied. To provide a high impedance connection between the capacitor and the sample and hold amplifier, a glass-encased Form A reed relay and high voltage wiring were used. To provide the separate storage of two analyses, two sets of reed relays and sample and hold amplifiers were used.

Selection of the proper anlaysis signal storage must be synchronized with the last sample injected into the analyzer. During automatic switching from one stream to the other, selection of the correct reed relay was made through contacts on the same relay that was next to be analyzed. This was accomplished through the use of a double-pole double-throw ratchet relay (which reverses the sense of the relay contacts only upon energizing). If single stream operation was desired, the sense of the reed relay selection contacts were reverse and further operation of the ratchet relay inhibited. This was accomplished through a triple-pole double-throw switch that opened the ratchet relay coil circuit as it reversed the reed relay coil circuits. An electrical schematic of the controls is shown in Figure 12.

The three-position toggle switch located on the side of the peak picker and memory control enclosure provided three modes of sample injection sequence. Two indicator lights, identified as stream 1 and 2, displayed which stream was next to be analyzed. Stream 1 was the feed stream; stream 2 was the homogenized aerator sample. When the switch was in the AUTOMATIC position, the controls caused alternate injection and memory of the two samples. In the MANUAL position, only the feed stream was analyzed with its associated memory updated. In the STANDARD position, the memory update function was inhibited so that analyzer calibration procedures did not affect the stored analysis values for the two streams.

SHEET NO

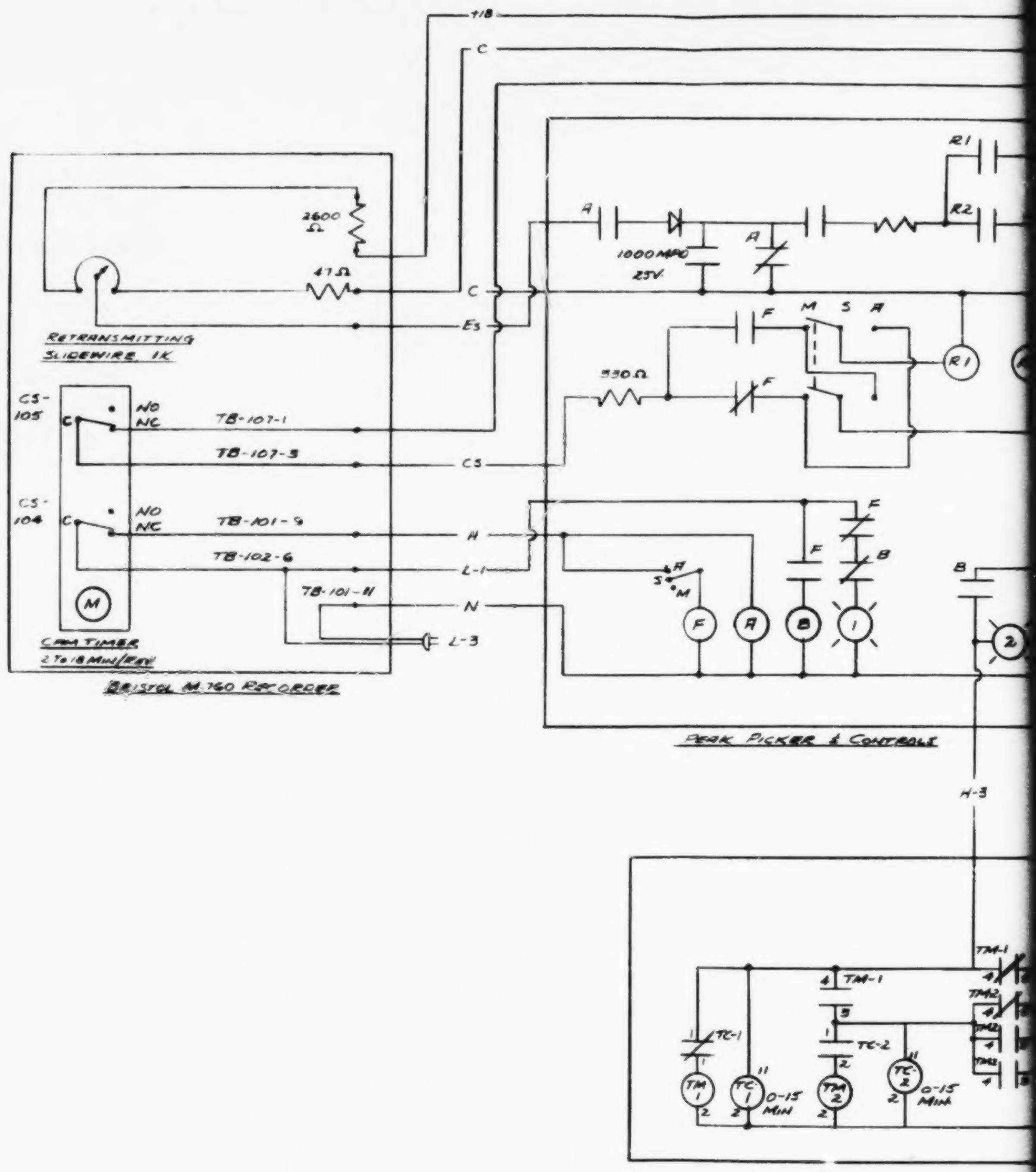


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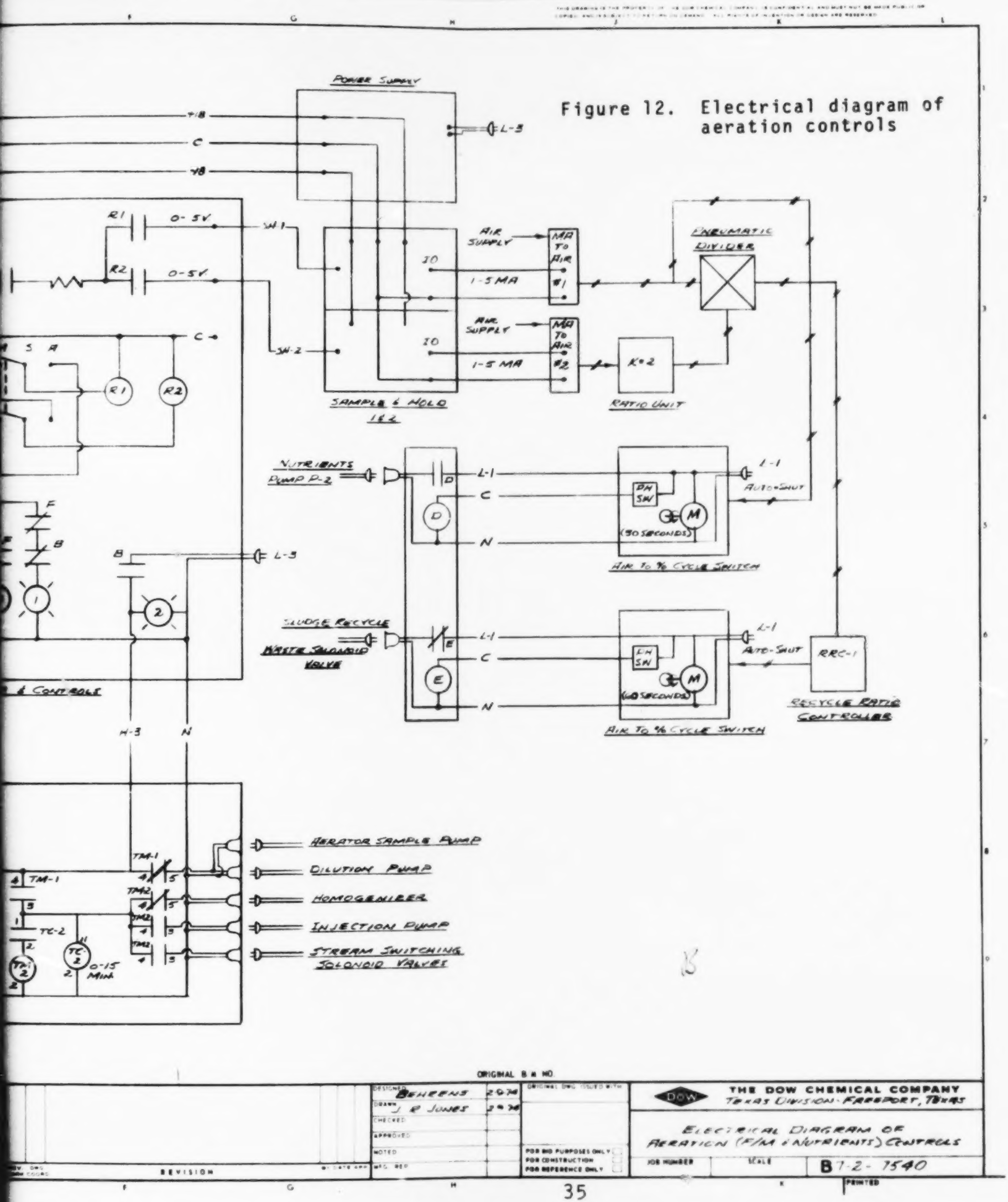
Figure 12. Electrical diagram of aeration controls



F/M Signal Conversion and Control

A block diagram of the food to microorganisms ratio control equipment is shown in Figure 13 (the nutrients control is also included in this diagram).

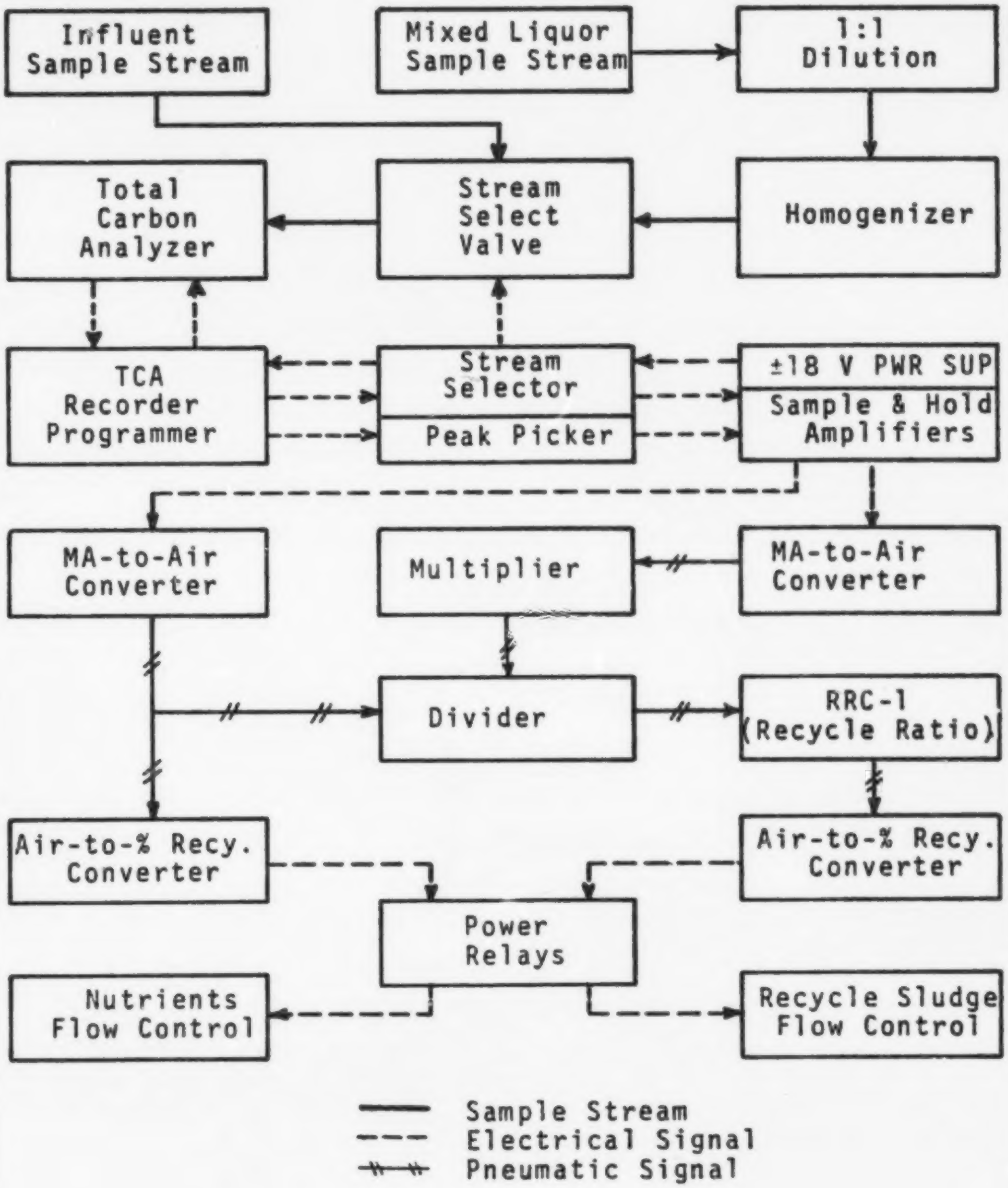
The output of each sample and hold amplifier was converted to a 1- to 5-ma. signal and then to a 3- to 15-psi pneumatic signal. Since the biota sample had been diluted 1:1, the value of that signal was multiplied by two before division into the signal for the food. The quotient of the pneumatic divider was the F/M ratio with a 0 to 1 range. This signal was fed to a pneumatic recorder-controller. The controller adjusted an output pneumatic signal proportional to the offset between the desired and actual F/M value. This output signal controlled the percent energized time of a 60-second-cycle timer. The energized circuit operated a one-half inch NPT normally closed solenoid valve that allowed flow of the bio-settler underflow stream to waste. This stream was pumped at constant flow and was normally recycled to the aeration basin inlet.

The system also included a high signal selector and regulator that can be set to designate a maximum waste flow rate independent of the F/M controller. This safety device prevented cell washout in case of control system failure.

THEORETICAL BACKGROUND AND BASIS FOR UNSTEADY-STATE COMPUTER SIMULATION

Unsteady-state material balances coupled with reaction kinetics were the basis for the development of a computer simulation of biological growth, substrate removal, and the F/M control system dynamics.

Figure 13. Nutrients and F/M control---block diagram



Growth and Substrate Utilization Kinetics

The basic emperically developed kinetic relationship generally used to describe the growth rate of microorganisms in a biological oxidation system is:

$$\frac{dX}{dt} = \frac{YdF}{dt} - k_d X \quad (1)$$

where X = concentration of microorganisms (mass/volume)

$\frac{dX}{dt}$ = net growth rate of microorganisms (mass/volume-time)

Y = growth yield coefficient (mass of microorganisms produced/mass of substrate utilized)

$\frac{dF}{dt}$ = rate of substrate utilization (mass/volume-time)

k_d = microorganism decay coefficient (time^{-1})

Another emperical relationship, similar to that developed by Monod (1949), can be used to describe the rate of substrate utilization by microorganisms:

$$\frac{dF}{dt} = \frac{kXS_1}{K_s+S_1} \quad (2)$$

where k = maximum rate of substrate utilization per unit weight of microorganisms (time^{-1})

K_s = half velocity coefficient, equal to the substrate concentration when $(dF/dt)/X = (1/2)k$ (mass/volume)

S_1 = concentration of substrate surrounding the microorganisms (only that portion of the waste which is soluble and biodegradable) (mass/volume)

Division of both sides of Equation (2) by X gives

$$\frac{dF/dt}{X} = \frac{kS_1}{K_S+S_1} \quad (3)$$

Hence, the removal of substrate per unit mass of microorganisms on a finite time basis defined as $(F/M)_r$ is given by:

$$\frac{\Delta F/\Delta t}{X_{avg.}} = \frac{kS_1}{K_S+S_1} \equiv (F/M)_r \quad (4)$$

where X_{avg} is the average biota concentration present in the aeration basin during the chosen time interval.

The process loading ratio (F/M) is defined in terms of the efficiency of removal as:

$$F/M \equiv \frac{(F/M)_r}{E} \times 100 \quad (5)$$

where E = $100\,(S_0-S_1)/S_0$ (percent efficiency)
S_0 = influent waste concentration (soluble substrate) (mass/volume)

Equations (4) and (5) implicitly illustrate an advantage of F/M control. If F/M is held constant for a designated treatment efficiency, then the removal per unit mass $(F/M)_r$, is fixed. By Equation (4), constant $(F/M)_r$ implied a constant effluent soluble waste concentration, S_1, since k and K_S are constants. In other words, if the amount of substrate available to a unit mass of microorganisms remains constant then the degradation by the biota is fixed.

Another parameter commonly calculated in the activated sludge treatment is the sludge age. Writing Equation (1) on a finite time basis after division by X yields

$$\mu = \frac{\Delta X/\Delta t}{X_{avg}} = Y(F/M)_r - k_d \qquad (6)$$

where μ is the net specific growth rate of microorganisms. The reciprocal of μ is what is termed as the sludge age, or mean cell residence time and is inversely proportional to F/M.

Unsteady-State Modeling of the Activated Sludge Process

The Monod kinetic equations can be incorporated into substrate and microorganism material balances around the aeration basin (Figure 14) and provide a basis for mathematical modeling of the activated sludge process. The substrate material balance is:

$$QS_0 + RS_1 - (Q+R)S_1 - \frac{kS_1XV}{K_s+S_1} = V\frac{dS_1}{dt} \qquad (7)$$

or

$$Q(S_0-S_1) - \frac{kS_1XV}{K_s+S_1} = V\frac{dS_1}{dt} \qquad (8)$$

and the microorganism material balance is:

$$RX_r + V\left(\frac{YkS_1X}{K_s+S_1} - k_dX\right) - (Q+R)X = V\frac{dX}{dt} \qquad (9)$$

where Q = influent waste flow (volume/time)

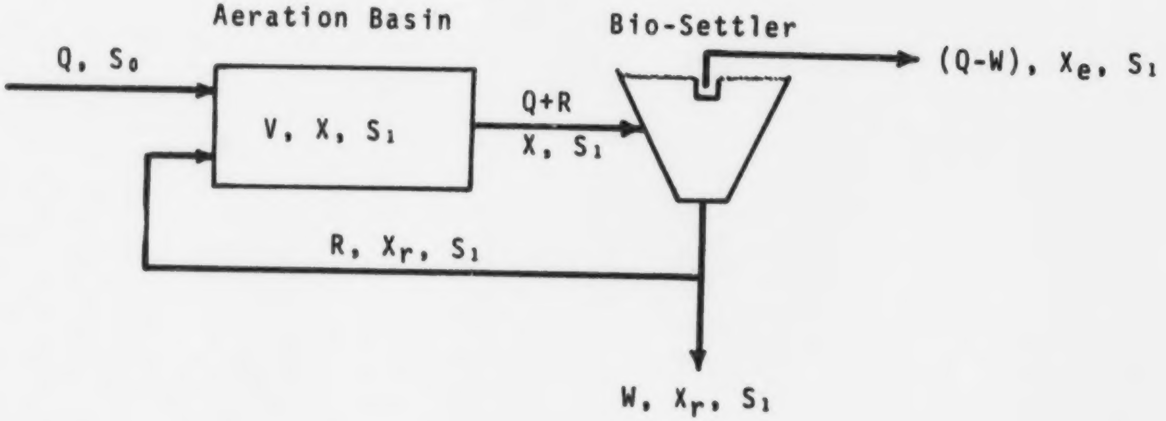
R = recycle sludge flow (volume/time)

W = waste sludge flow (volume/time)

V = aeration basin (volume)

X_r = bio-settler underflow sludge concentration (mass/volume)

Figure 14. Schematic of an activated sludge process completely mixed biological reactor



LEGEND

Q	Influent Waste Flow, volume/time
R	Recycle Sludge Flow, volume/time
W	Waste Sludge Flow, volume/time
V	Aeration Basin Volume
S_0	Influent Substrate (waste) Concentration, volume/time
S_1	Reactor Substrate Concentration, mass/volume
X	Reactor Microorganism Concentration, mass/volume
X_r	Sludge Recycle Concentration, mass/volume
X_e	Effluent Microorganism Concentration, mass/volume

Implicit for this model is the assumption of complete mixing in the aeration tank. The model is conservative in that substrate degradation is assumed to occur only in the aerated reactor.

The scheme of F/M proportional control can easily be inserted into the activated sludge model. The waste flow is varied with F/M by:

$$W = A(F/M) + B , \quad 0 \leq W \leq W_{max} \tag{10}$$

where A and B are constants relating to a specified gain. A maximum waste rate, W_{max}, is usually imposed to prevent complete cell washout in case the control system should fail. The control is related to the recycle flow by

$$T = R + W \tag{11}$$

$$R = T - [A(F/M) + B] \tag{12}$$

where T is the total bio-settler underflow rate.

A digital computer program was developed to model the transient biological state during F/M control and is presented in Appendix C. A fourth order Runge Kutta method using Gill's coefficients (Gill, 1951) was used to numerically integrate the material balance differential equations and Equation (12) was used for recycle control.

The dynamics of the unsteady-state bio-settler operation are computed in terms of a constant compaction ratio, X_r/X.

$$\frac{X_r}{X} = \frac{n}{100} \frac{Q+R}{R+W} \tag{13}$$

where η is defined as the efficiency of settler operation. This equation can be derived from a steady-state material balance around the bio-settler, assuming that a negligible mass of microorganisms leaves through the clarifier overflow (η = 100 percent).

Figure 15 shows the fit of Equation (13) for steady-state miniplant data, indicating a settling efficiency of 97 percent.

There were obvious limitations in using a constant ratio to describe the non-steady-state performance of a biological settler. However, no better general model could be found and this relationship was able to adequately allow predictions of unsteady-state bacteria concentrations in the aeration basin. Figure 17 shows that the sludge compaction ratio did remain relatively constant during unsteady-state miniplant operation.

It should be pointed out that having a constant sludge compaction ratio was specific for a particular settler geometry and system. This consideration was particularly important in scale-up design. Tracey and Keinatch (1973) present a dynamic model based on gravitational and bulk flow considerations which does not indicate a constant sludge compaction ratio. Their model predicts an increase in sludge blanket height but no change in sludge underflow concentration for an increased solids input.

KINETIC CONSTANTS FOR PROPYLENE GLYCOL ACCLIMATED BIOTA

The four kinetic constants represented in Equations (1) and (2) were determined from miniplant TOD and MLVSS data. The data were retrieved from aeration basin temperature at 30 ±3° C.

Figure 15. Bio-settler steady-state compaction fresh water glycol bacteria

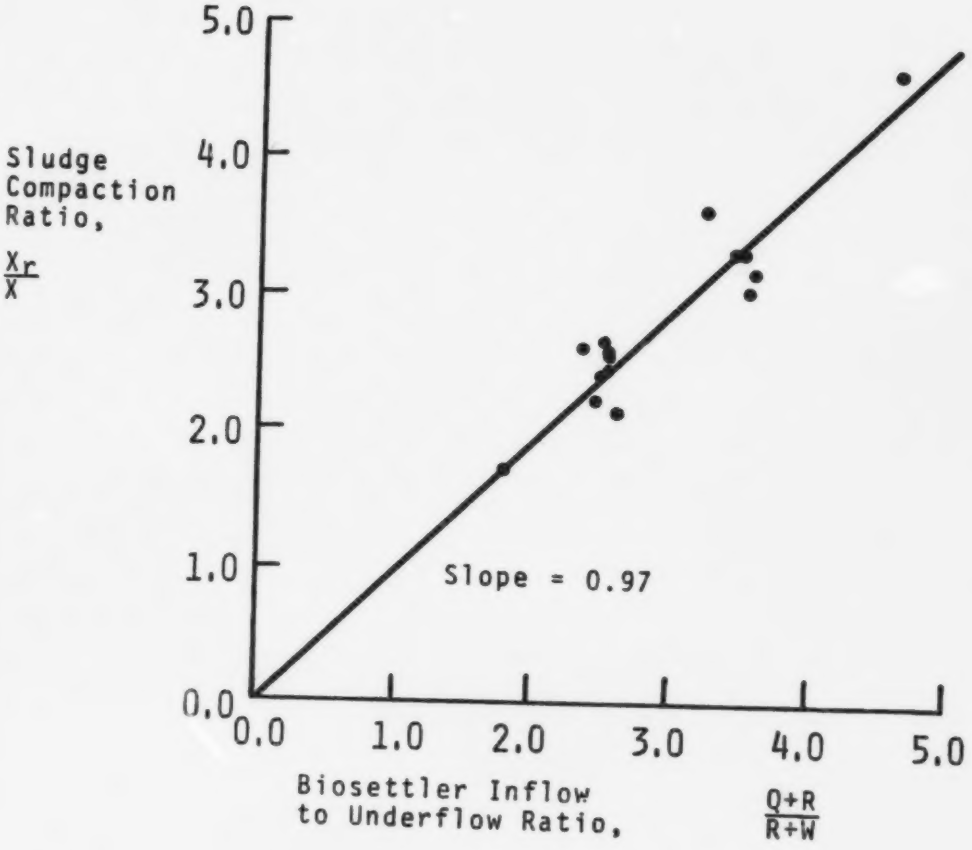
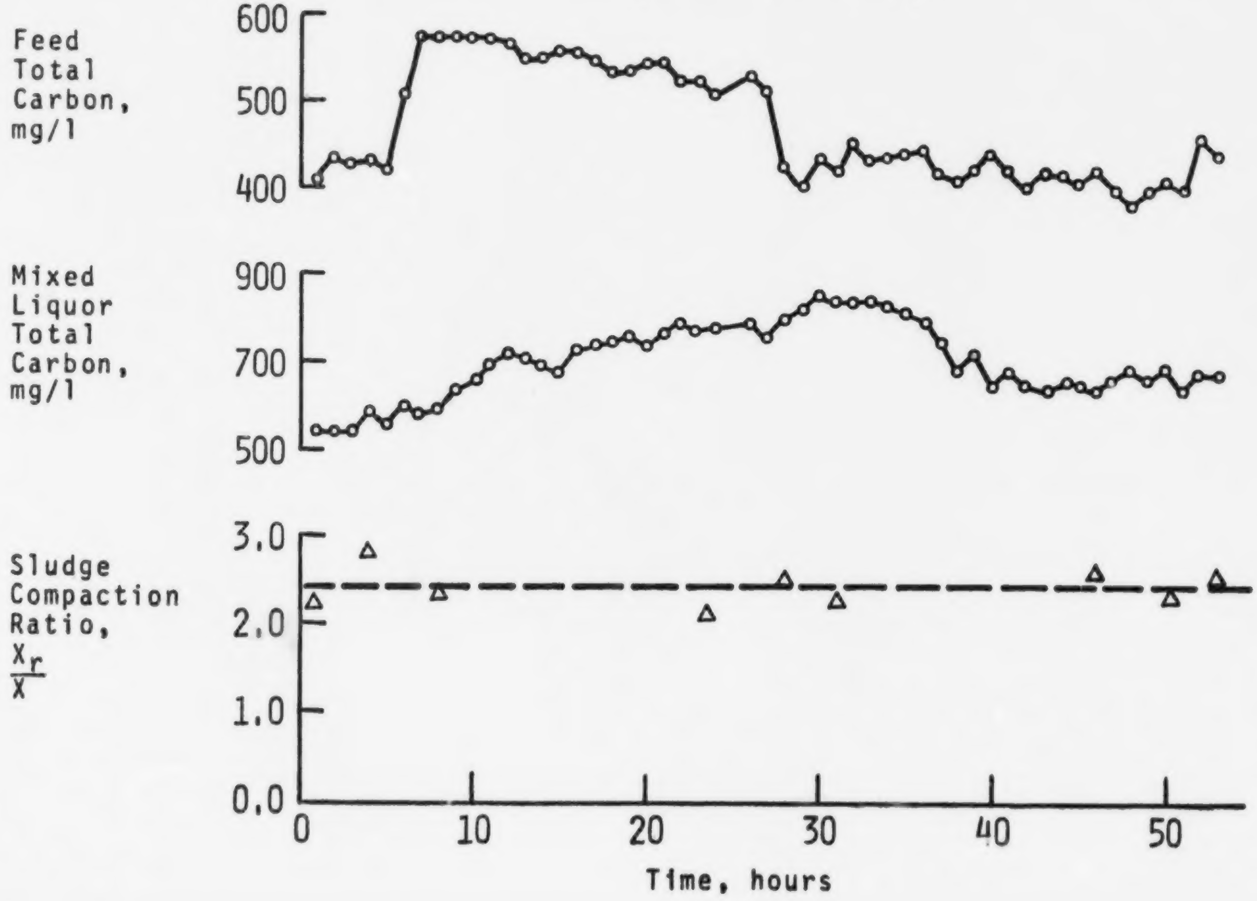


Figure 16. Unsteady-state variation of the sludge compaction ratio (data from run C)



The analysis for TOD feed data was done on a sample which had been passed through a 3.0-micron filter. TOD bio-settler overflow data was used as the representation of soluble substrate in the aeration basin, S_1, and a sample which had been passed through a 25-micron filter was analyzed.

Figure 17 shows the determination of the sludge growth coefficients for the fresh water propylene glycol biota. The slope of the line is the yield coefficient ($Y = 0.21$ gm VSS formed/gm TOD removed) and the ordinate intercept is the decay or endogenous coefficient ($k_d = 0.06$ day^{-1}). The curve is fitted by a least-squares analysis and the coefficient of correlation is 0.81.

The top graph in Figure 18 shows the substrate utilization per unit mass of microorganisms as a function of the TOD (soluble substrate) of the bio-settler overflow. In order to accuarately obtain the fresh water substrate removal kinetic constants, Equation (4) can be linearized by using the reciprocal equation (Gaudy, Ramanathan, and Rao, 1967).

$$\frac{1}{(F/M)_r} = \frac{K_s}{kS_1} + \frac{1}{k} \qquad (14)$$

The reciprocal data, plotted in the lower graph of Figure 18 is fitted by a least-squares anlaysis and the coefficient of correlation is 0.84. The value of k is 1/intercept, $k = 1/0.2 = 5.0$ day^{-1} and the value of K_s is k times the slope, $K_s = 5.0 \times 23.5 = 117$ mg/l TOD.

For the salt water bacteria, K_s is large in comparison to S_1 and Equation (2) reduces to:

$$dF/dt = k'XS_1 \qquad (15)$$

Figure 17. Sludge growth coefficients---fresh water-glycol feed

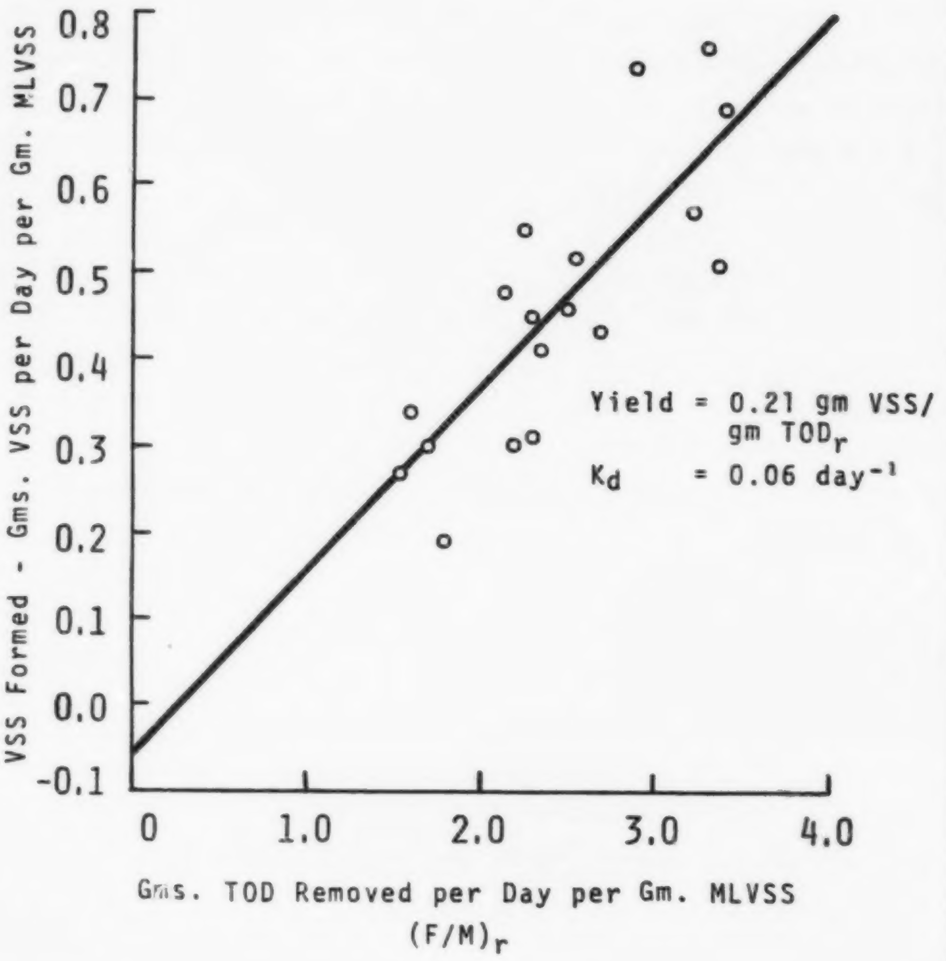
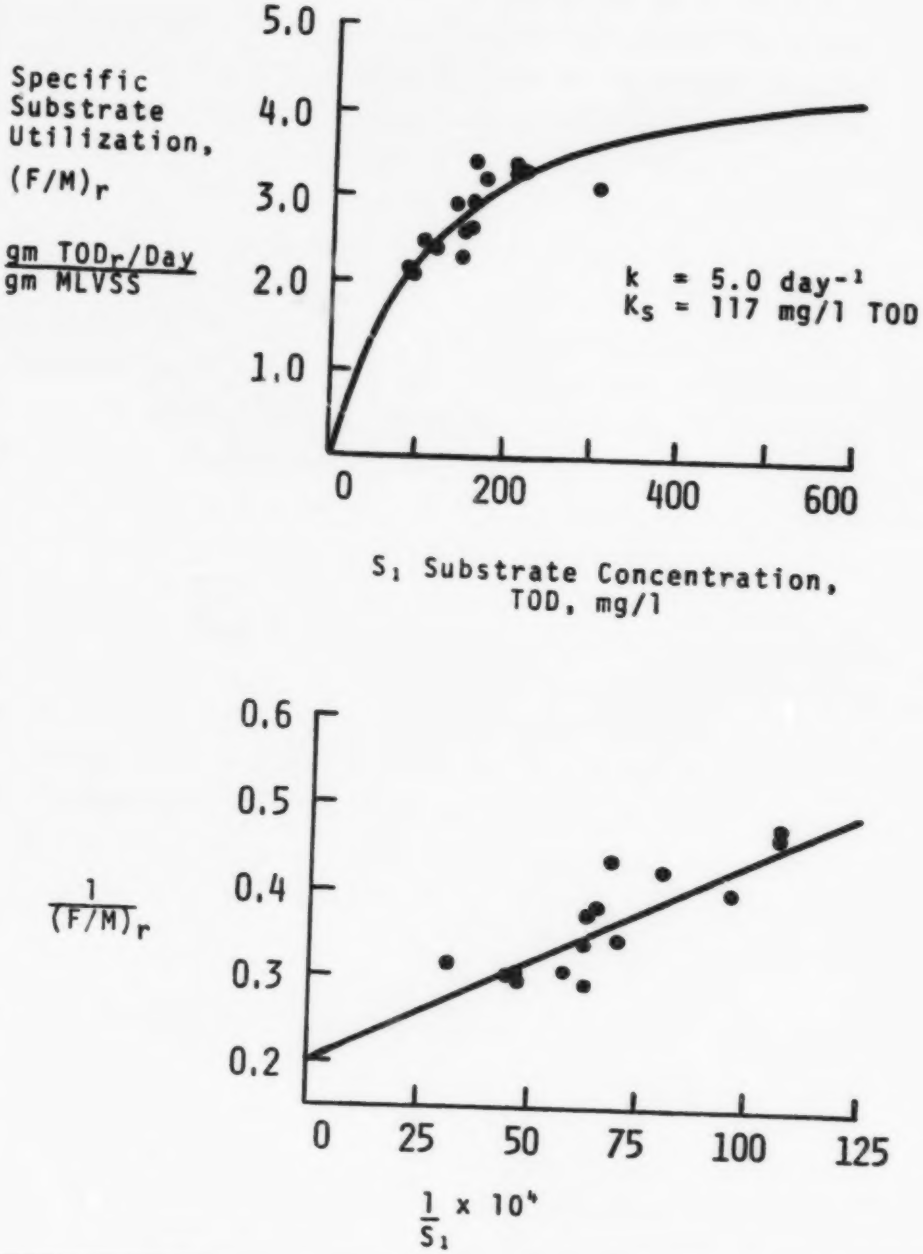


Figure 18. Substrate removal coefficients---fresh water-glycol feed



The kinetic constants were determined from data of a previous study on the treatment of the brine propylene glycol waste water (Zeitoun, 1971) as $k' = 0.0132$ (mg/l TOD)$^{-1}$day^{-1}, $Y = 0.1$-3 gm VSS formed/gm TOD removed, and $k_d = 0.02$ day^{-1}.

There has been much discussion in the literature on the limitations of using the Monod type kinetic equations to describe the dynamics of cell growth and substrate removal in a continuous flow biological oxidation system. Assumptions of a constant yield and decay coefficient have been criticized (Chiu, Fan, Kao and Erickson, 1972; Storer and Gaudy, 1969; McLellan and Busch, 1969) and it has been suggested that K_s be considered dependent upon the degree of mixing (Kornegay and Andrews, 1970). Changes in viability and physiological activity of the microorganisms relate to the difficulty in obtaining descriptive kinetic coefficients. However, a degree of success can be anticipated in using this model to simulate the dynamics of the propylene glycol-activated sludge miniplant because of four basic reasons:

1. The Monod model was originally developed to describe behavior of pure cultures. It was found that about 80 percent of the microorganisms present in the fresh water glycol system were the one bacteria species, *Pseudomonas*. No more than two to four different bacteria are present in the brine-activates sludge and about 90 percent are one bacterium (Zeitoun, et al. 1971).

2. S_1 was in terms of a single, soluble substrate, propylene glycol.

3. The carbon energy source was rate limiting. Otherwise deficient nutrients were added by the nutrient control system.

4. The pH and temperature were regulated for the optimum rate of biological growth.

TESTING OF THE F/M CONTROL SYSTEM

In general, the miniplant test runs were started with the F/M controller set on manual, i.e., constant recycle and sludge waste flow rates were set. The system was allowed to approach a steady-state condition in terms of reactor microorganism concentration, substrate removal efficiency, and bio-settler characteristics such as sludge concentration and sludge blanket height. The system was then put on automatic control and a step increase in the concentration of the feed was made by adding an aliquote of industrial grade propylene glycol to the equalization tank. The aeration basin temperature was held constant during each run.

An automatic, total carbon analysis was made every six minutes during these runs, with alternate analysis of feed sample and homogenized mixed liquor sample. To reduce the volume of data, each F and M concentration plotted in this section is represented as the average of five total carbon analyses. Waste, recycle, and F/M data are also hourly averaged values.

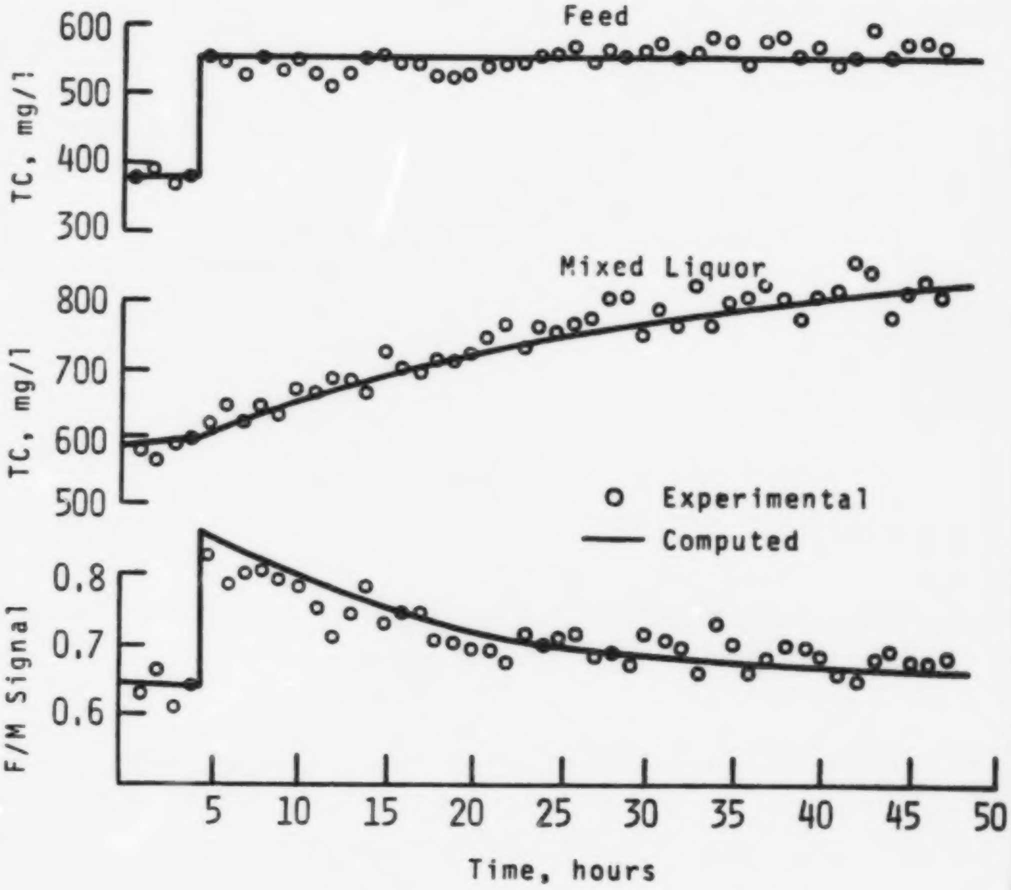
A blank F/M test in which no control was effected, that is, the recycle and waste flow rates were kept constant throughout the test, showed the dynamics of cell growth under continuous flow conditions and was used as a basis for comparison for F/M control test system responses.

Fresh Water Propylene Glycol System

Figure 19 shows the mixed liquor and F/M response to a 47-percent increase in loading from Blank Run A. The solid

Figure 19. F/M blank test---
(Constant Recycle Flow, Run A)

Feed Flow	1938 ml/min
Sludge Recycle Flow	1070 ml/min
Sludge Waste Flow	85 ml/min



lines are the predicted responses using the transient state computer model (Appenix C). A very good fit was obtained for the increase in MLTC (mixed liquor total carbon) but, because of a non-linear calibration of the total carbon instrument above 50 percent of scale, data for the F/M signal lay below the computer prediction of F/M. When the MLTC values reach the same non-linear range as the feed values, the predicted curve agreed with the data.

Analytical technique did not entail separate determination of the microorganisms and substrate concentrations in the aeration basin. However, the concentration of substrate in the reactor represented less than five percent of the MLTC concentration under steady-state conditions. The computer simulation shown in Figure 20 illustrates the effect of the step increase in loading upon the reactor substrate concentration during Blank Run A. The substrate concentration rapidly increased after the step change but the dynamics of cellular growth caused the food to microorganisms ratio to steadily decrease, thus causing the substrate concentration in the aeration basin to gradually decrease to the original steady-state level.

In Table 3 and Figure 21 the F/M testing conditions for Blank Test A and four F/M control tests are summarized. The initial experimental conditions for Control Test B (Figure 22) were similar to those for Blank Test A. However, the initial mixed liquor concentration in Run B was 120 mg/l higher than that of Run A, even though the initial feed concentration only differed by 25 mg/l.

Computer simulation verified that the bacteria concentration in Run A had not yet reached a steady-state level at the

Figure 20. Relationship between the mixed liquor, microorganisms, and substrate concentrations after a step increase in loading---Blank Run A, computer simulation

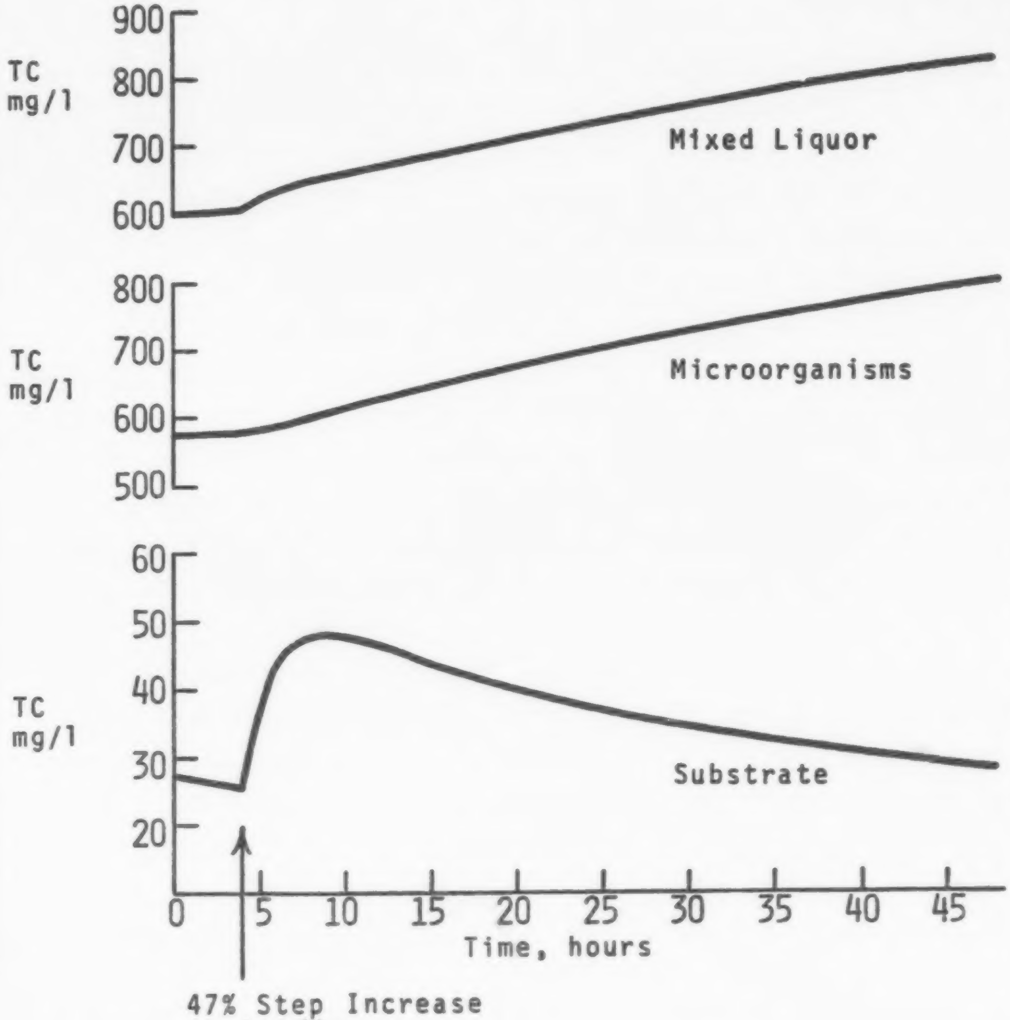


Table 3. F/M TEST CONDITIONS

Run	Constant Conditions		Initial Conditions				F/M Set Point
			Concentrations		Flow Rates		
	Feed Flow, ml/min	Aeration temp. °C	Feed, mg/l TC	Mix. Liq., mg/l TC	Recycle, ml/min	Waste, ml/min	
Blank[a] Test A	1938	29 ±1	385	600	1070	85	0.6[a]
Control Test B	1938	29 ±1	410	720	1070	85	0.6
Control Test C	1938	31 ±1	420	565	1060	85	0.6
Control Test D	2325	32 ±1	400	750	700	75	0.56
Control Test E	2325	33 ±1	515	800	975	85	0.56

[a] No control, set point determined by inherent stability where growth = wastage

initial feed concentration. About four days would have been required to reach steady state (Figure 21).

The computer simulation fitted the data of Run B (Figure 22, a&b). The drop in mixed liquor concentration from t=19 to t-28 hours was probably due to a plugged aeration basin outlet, causing a dilution of reactor mixed liquor by the feed stream. A rapid buildup of bacteria concentration occurred after the unit was unplugged because of an increased recycle sludge concentration.

Figure 22b demonstrates the proportionality involved in F/M control. Because of the increased recycle flow at the expense of sludge wastage, the F/M response of the control test was faster than the response for the blank.

The conditions of Run C (Figure 23, a&b) compared closely with those used in Blank Test A. The reactor microorganism concentration was initially below its steady-state level (Figure 23a) since the step decrease in feed concentration at t-27.5 hours, the original concentration, allowed the biota to proceed toward a steady-state level.

The model prediction of the increase in mixed liquor total carbon (t=0 to t=27.5 hours) fitted the data for Run C. A good representation of the recycle and wastage response is obtained with a fast F/M response (Figure 23b).

The F/M signal data lay above the predicted F/M curve for the first several hours following the step change because an average value for the feed concentration rather than the individual data points, was used in the simulation.

Figure 21. Computed initial steady state---for Run A

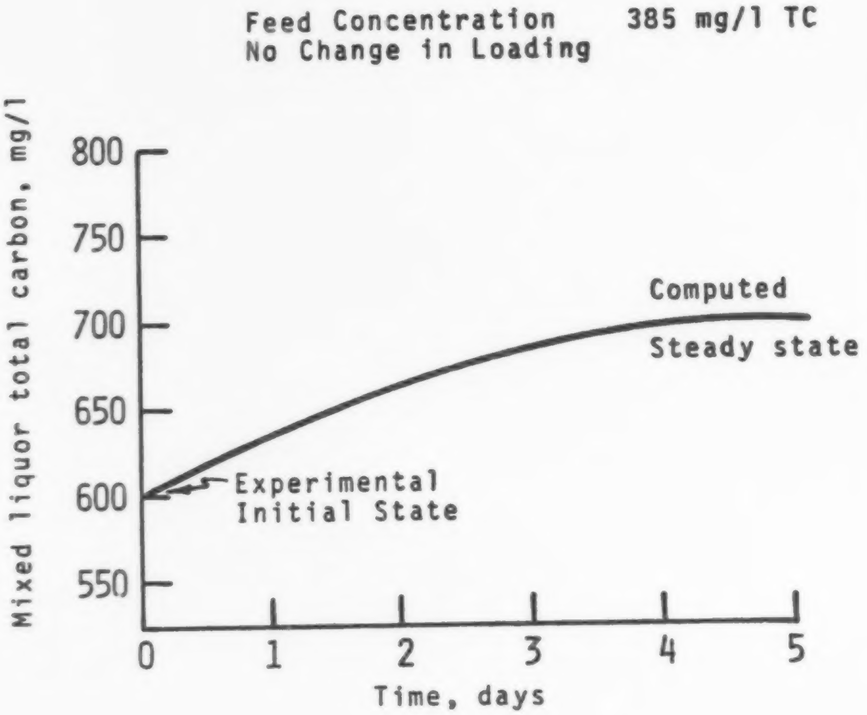
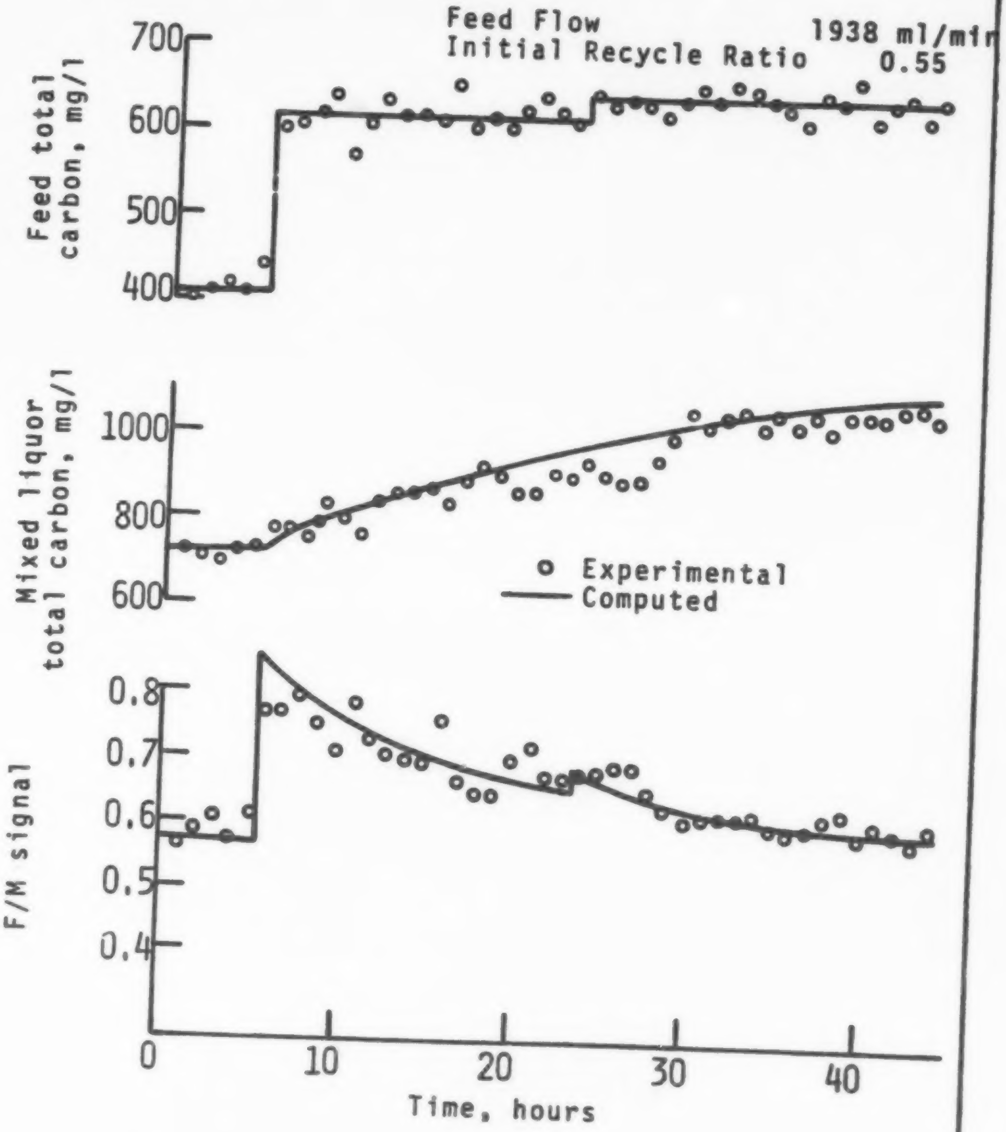


Figure 22a. F/M control test---
bacteria growth and F/M
response---Run B



Feed Flow 1938 ml/mi
Initial Recycle Ratio 0.55

Figure 22b. F/M control test---
recycle and waste response,
Run B

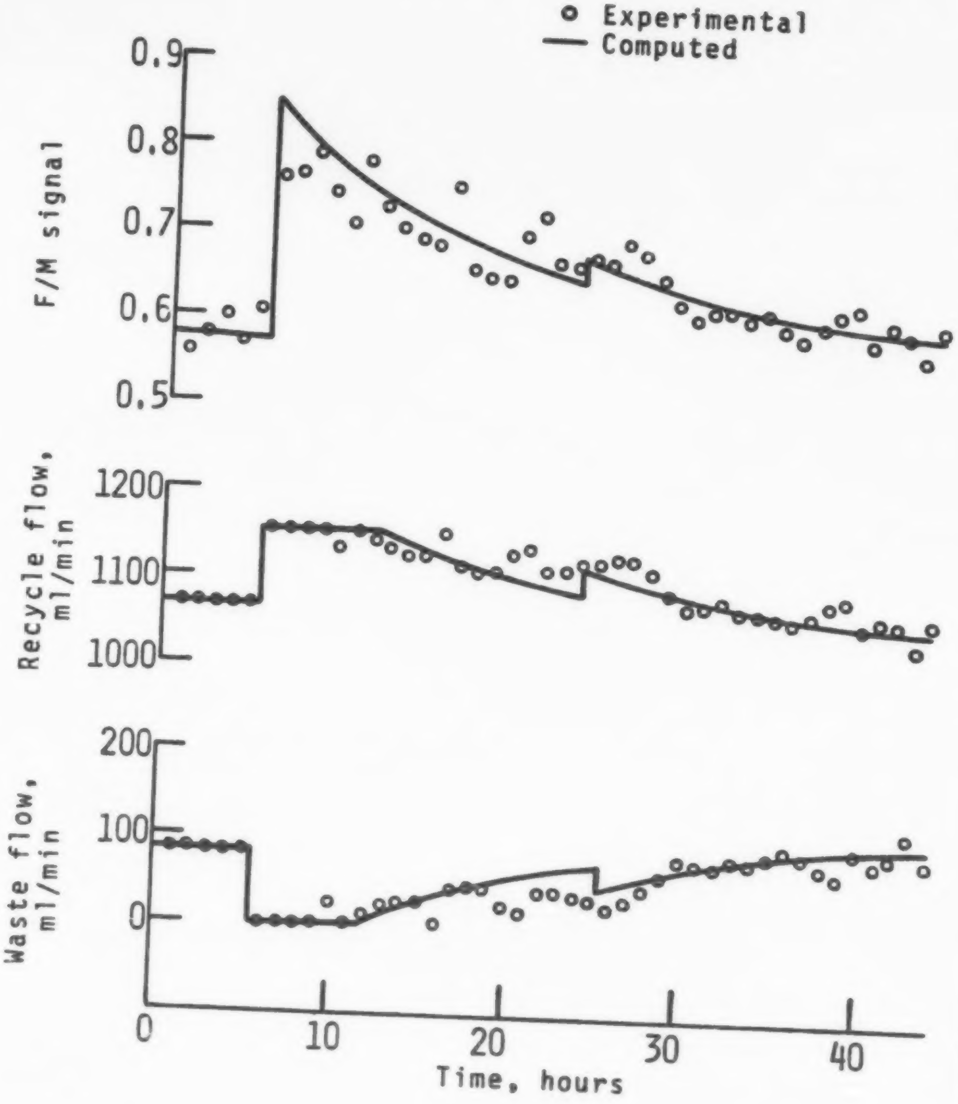


Figure 23a. F/M control test---
bacteria growth and F/M
response, Run C

Feed Flow 1938 ml/min
Initial Recycle Ratio 0.55

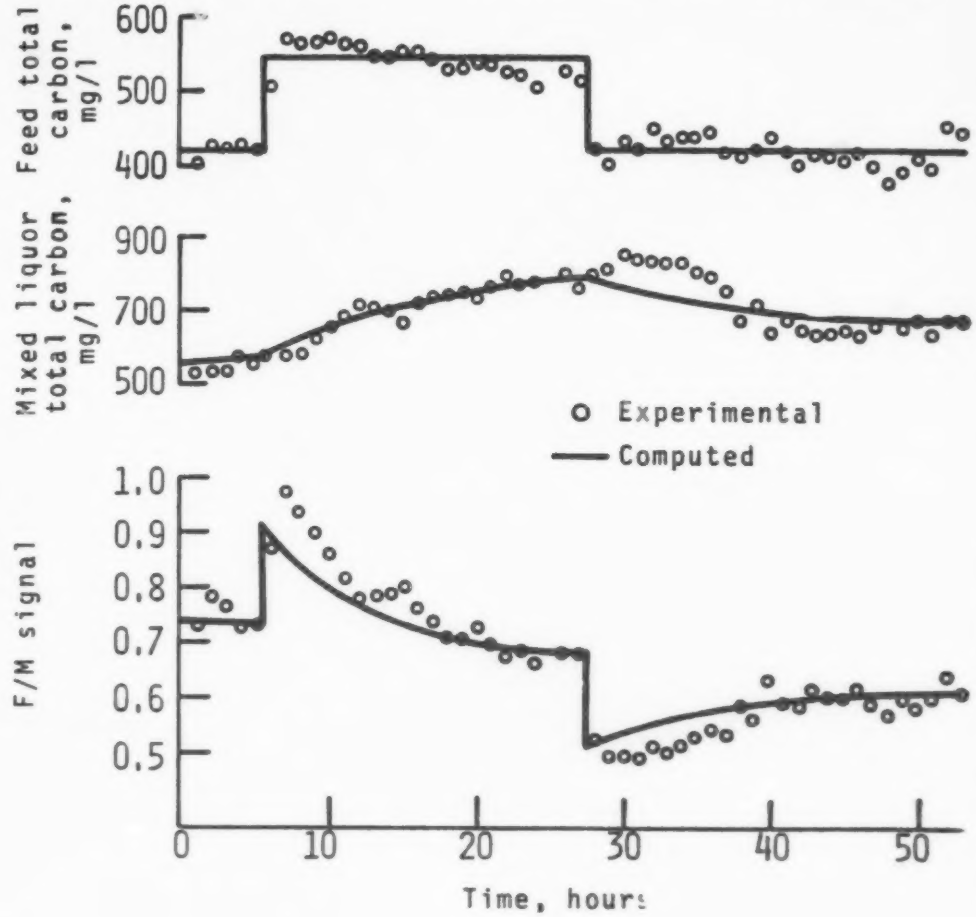
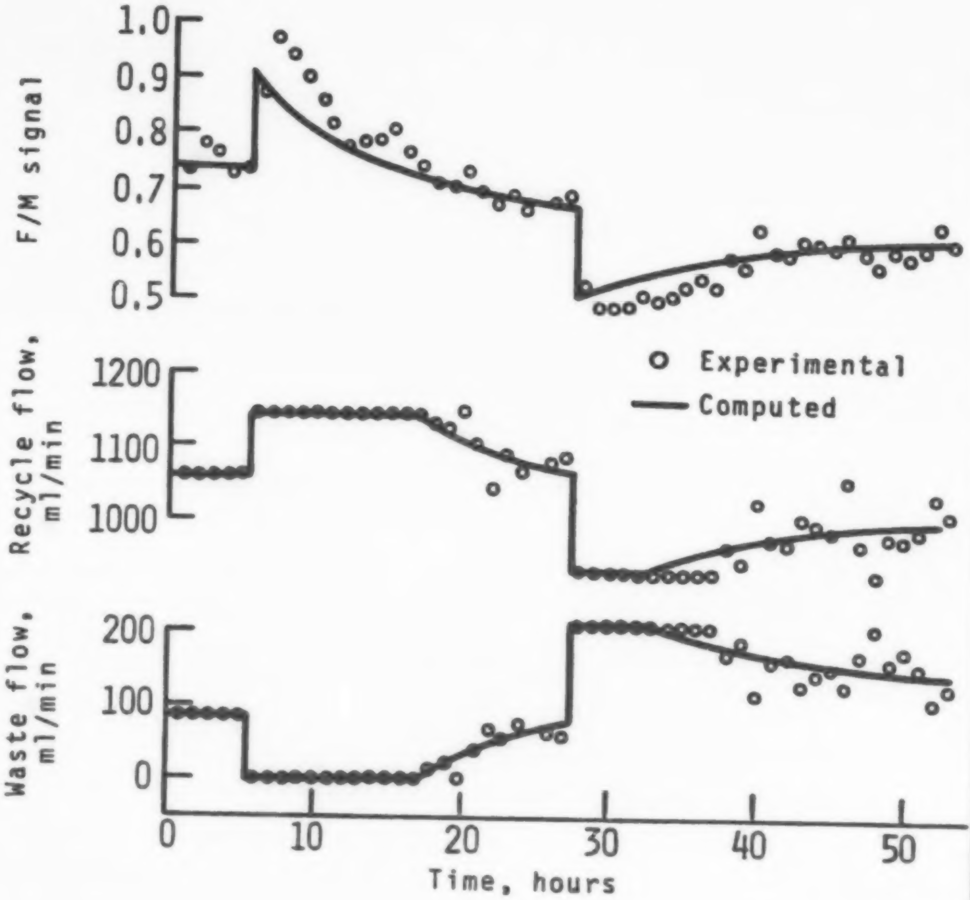


Figure 23b. F/M control test---
recycle and waste response,
Run C



The prediction of the transient biological state after the step decrease in feed concentration was not accurate because of an inadequate representation of the bio-settler dynamics. However, the final steady-state computer prediction (t > 50 hours) fitted the data.

Figures 24 a&b and 25 a&b show the results of two F/M tests (Run D and E) having different feed flows and recycle ratios. Increases in mixed liquor total carbon were well predicted by the mathematical model while the subsequent decreases were not accurately predicted. A more rapid return to the F/M setpoint was experienced in these runs because a greater mass flow of waste was available for additional recycle. Although the volumetric availability of waste was still 85 ml/min for Run E and only 75 ml/min for Run D, much higher concentrations of sludge existed because of higher compaction ratios (see Equation 13).

Figures 26 and 27 illustrate the effect that the F/M control had upon sludge settling. Figure 26 is plotted from early propylene glycol and MLVSS data. Abscissa values in Figure 27 are from on-line F/M control data, using the total carbon instrumentation.

A definite limited range of loading from 0.25 to 0.45 (permissible for good sludge settling) is shown in Figure 26. An optimum F/M of 0.35 gave a minimum SVI (sludge volume index) of less than 100. Operation at about 0.25 to 0.35 was desirable since a lower F/M also increased process efficiency in terms of soluble substrate (TOD) removal.

Figure 24a. F/M control test---
bacteria growth and F/M
response, Run D

Feed Flow 2325 ml/min
Initial Recycle Ratio 0.30

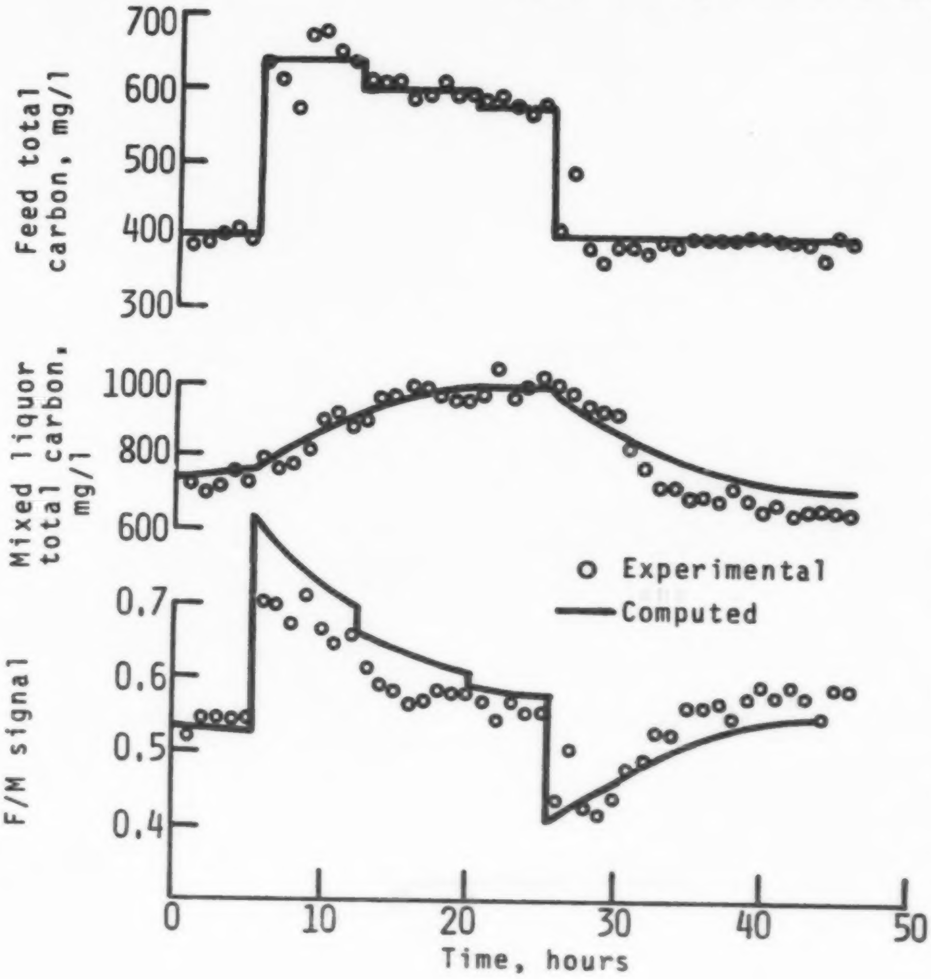


Figure 24b. F/M control test---
recycle and waste response,
Run D

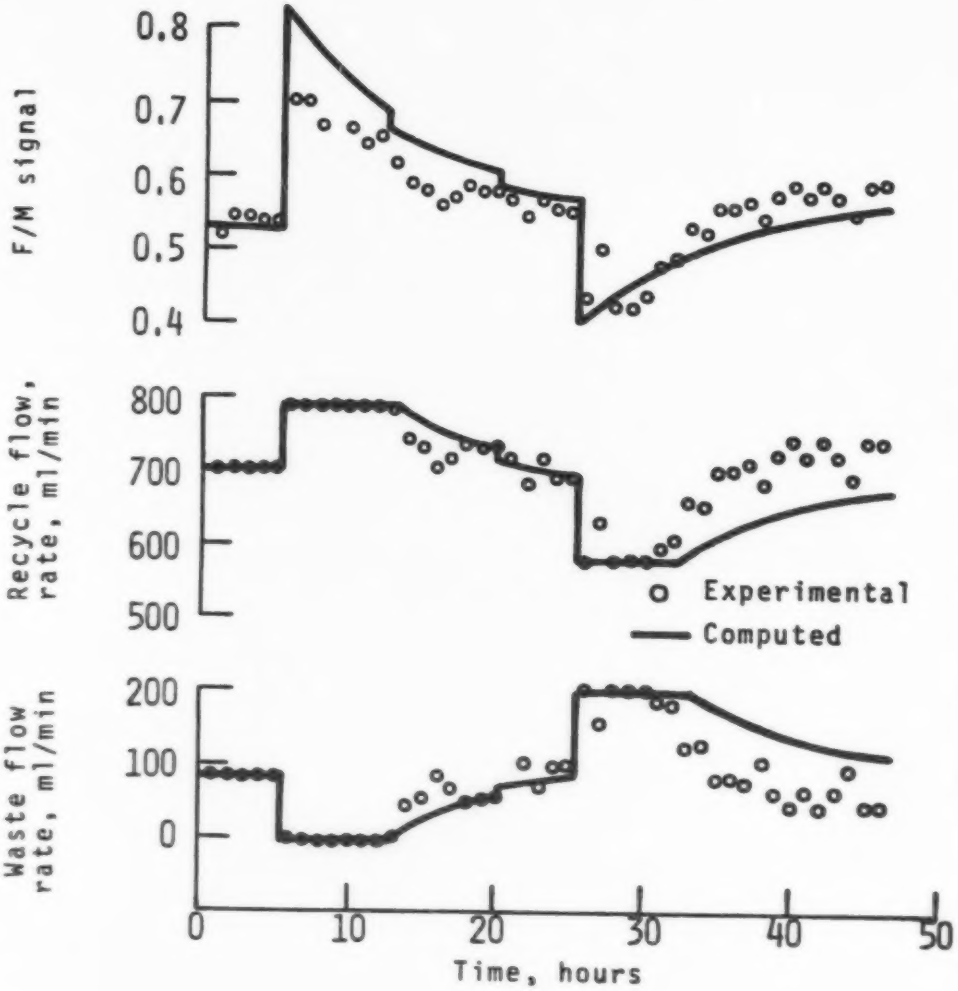


Figure 25a. F/M control test---
bacteria growth and
F/M response, Run E

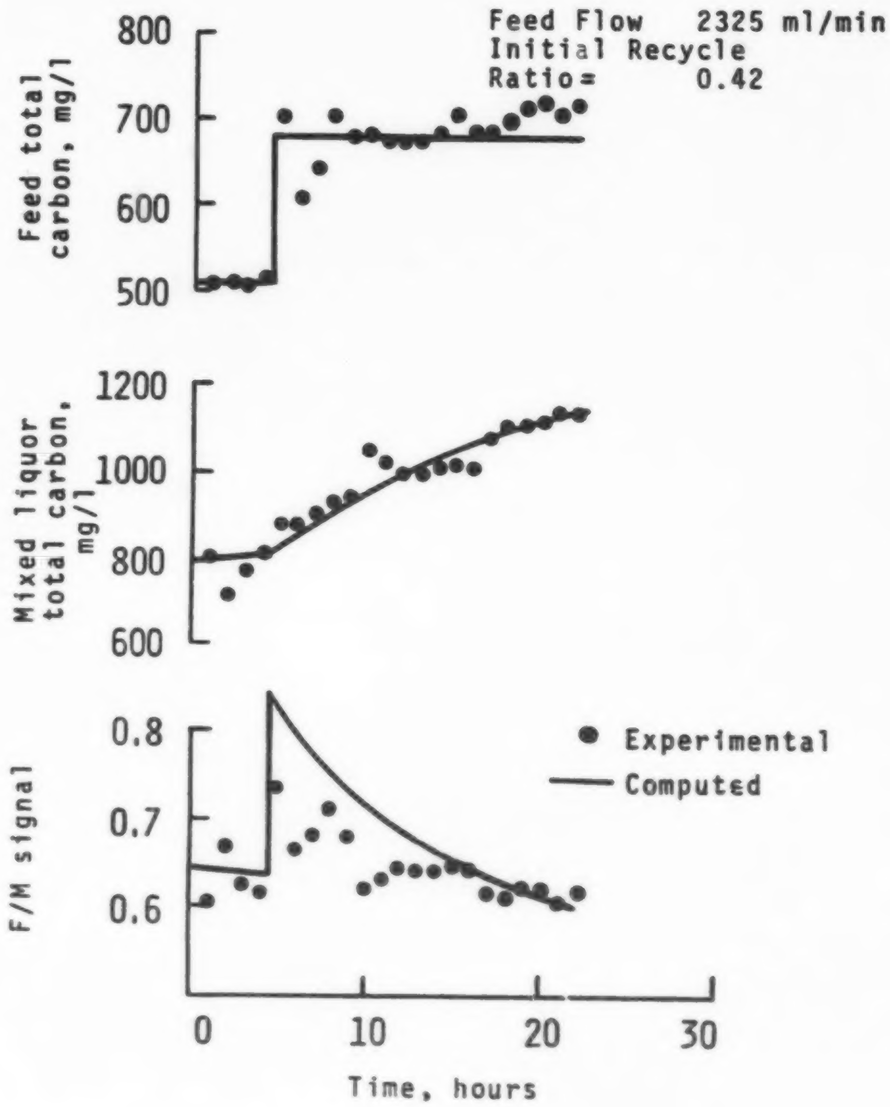


Figure 25b. F/M control test---recycle and waste response, Run E

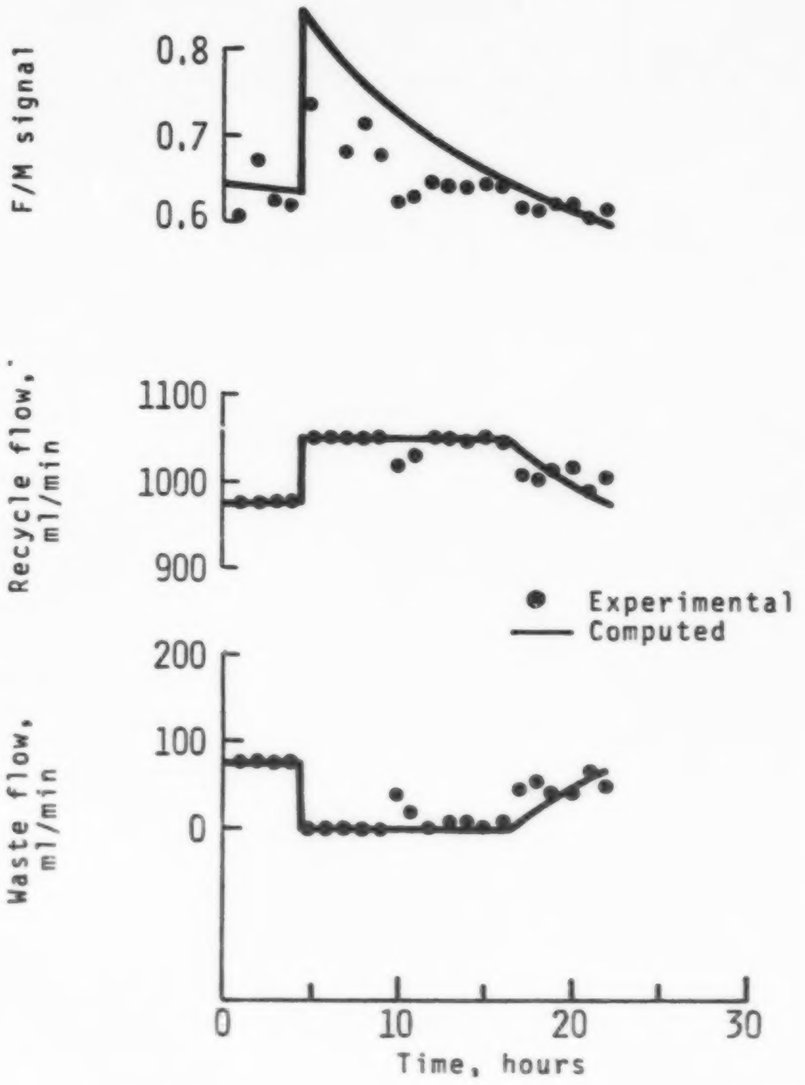


Figure 26. Sludge settling as a function of F/M (grab sample data)

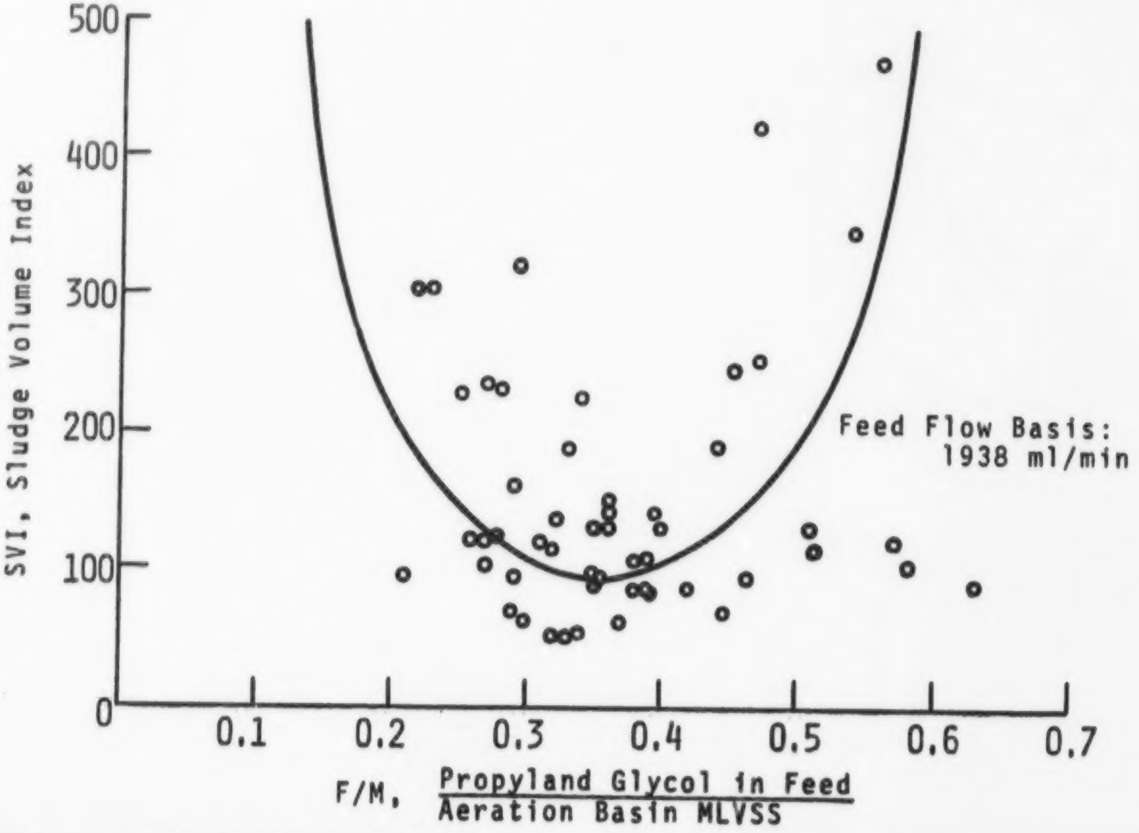
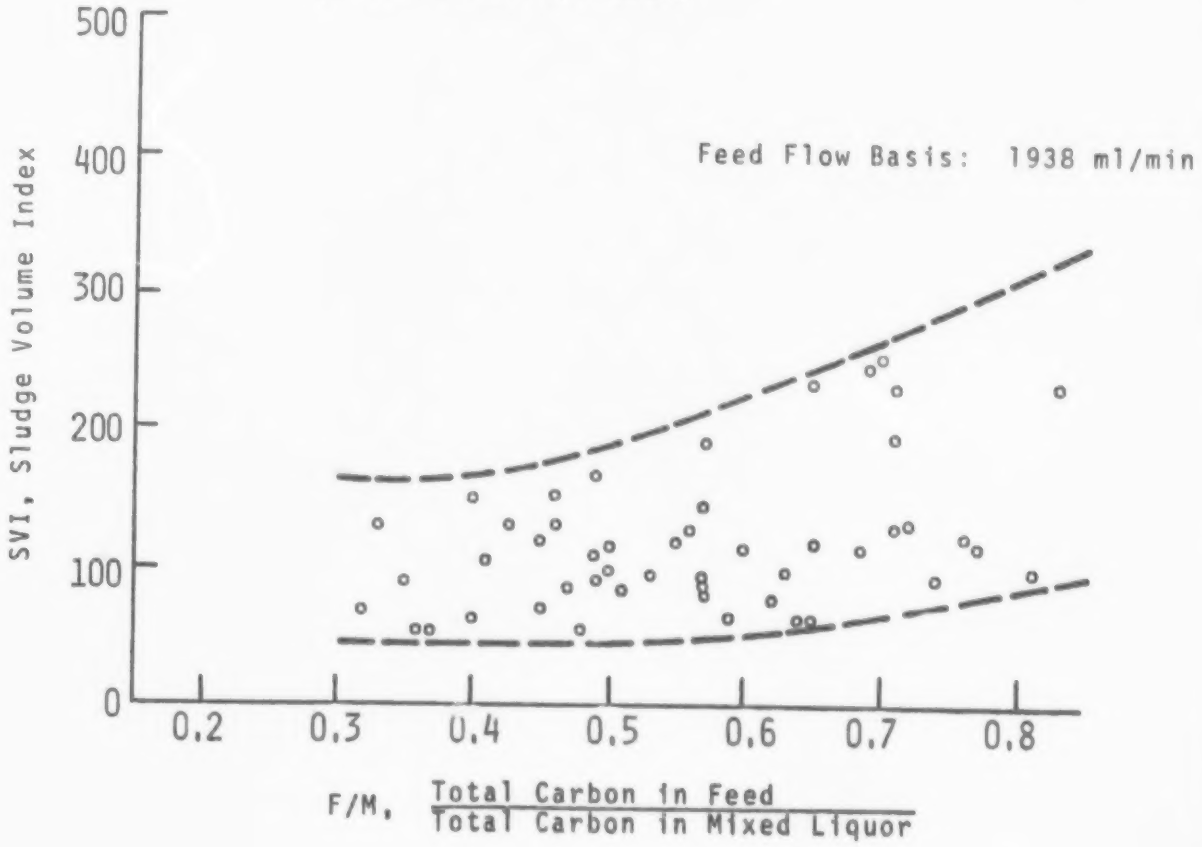


Figure 27. Sludge settling as a function of F/M (on-line control data)



The on-line control data in Figure 27 indicated an increase in the SVI at high loadings and an optimum F/M operating range from 0.3 to 0.5 TC_F/TC_M. With F/M, less than 0.3 increases in SVI were expected.

Saline Propylene-Glycol Waste Water System

The activated sludge miniplant was operated with the saline propylene-glycol waste water as the miniplant feed for about three weeks. During this period, the F/M control system did not result in a significant time reduction of F/M response because only 20 to 40 ml/min of waste sludge was available for additional recycle in the salt system. The growth rate of microorganisms was much smaller than in the fresh water system, as can be seen by inspection of the kinetic growth constants.

The waste water feed concentration upon start-up on the saline system and was only 250 mg/l total carbon. The initial biota concentration in the aeration basin was about 1000 mg/l MLVSS. There was no significant waste flow. The feed flow was 1938 ml/min (0.5 gpm), and the recycle ratio was about 0.6. Because of the low feed concentration (hence a very small growth of microorganisms) and the concentration loss of about 80 mg/l VSS in the bio-settler overflow stream, it was not possible to maintain the microorganism concentration in the aeration basin.

The feed concentration was gradually raised to about 450 mg/l total carbon by adding batch aliquots of propylene glycol to the equalization tank. With the feed concentration at 450 mg/l, the aeration basin MLVSS increased from 1082 to 2104 mg/l in three days (with no waste flow). The F/M control system served to maintain this level of biota concentration, but the response to increases in loading were slow.

SYSTEM EVALUATION

In order to have a meaningful basis of comparison for the F/M control tests, it was necessary to have a blank run with identical initial conditions. The initial steady state had to be the same in terms of feed, recycle and waste flow rates, substrate concentrations, reactor and bio-settler underflow microorganism concentration, temperature, and physiological conditions of the biota. Although these conditions were difficult to achieve experimentally, a valid comparison was made by using the mathematical modeling which predicted the transient accumulation of microorganisms.

The computer-simulated curves in Figures 28 and 29 show the differences in response characteristics for the fresh water propylene glycol system with F/M control versus the system with no control. The time-dependent simulations were allowed to proceed until a true steady state was reached before a 50-percent step increase in feed concentration was made.

The response time for F/M to return to the set point decreased markedly for the system with control. An 8.5 hour hydraulic residence time with an initial recycle ratio of 0.55 required 17 hours for the controlled system F/M to return within 20 percent of the set point level after the step increase in loading (Figure 28). This represented a 63-percent reduction in response time.

Figure 29 represents the system having a 6.95-hour hydraulic residence time and an initial recycle ratio of

0.3. The F/M ratio returned to within 20 percent of the set point in ten hours, less than a third of the time required for the system without control. The larger mass of waste available for recycle (as with Runs D and E) allowed a 69-percent reduction in response time.

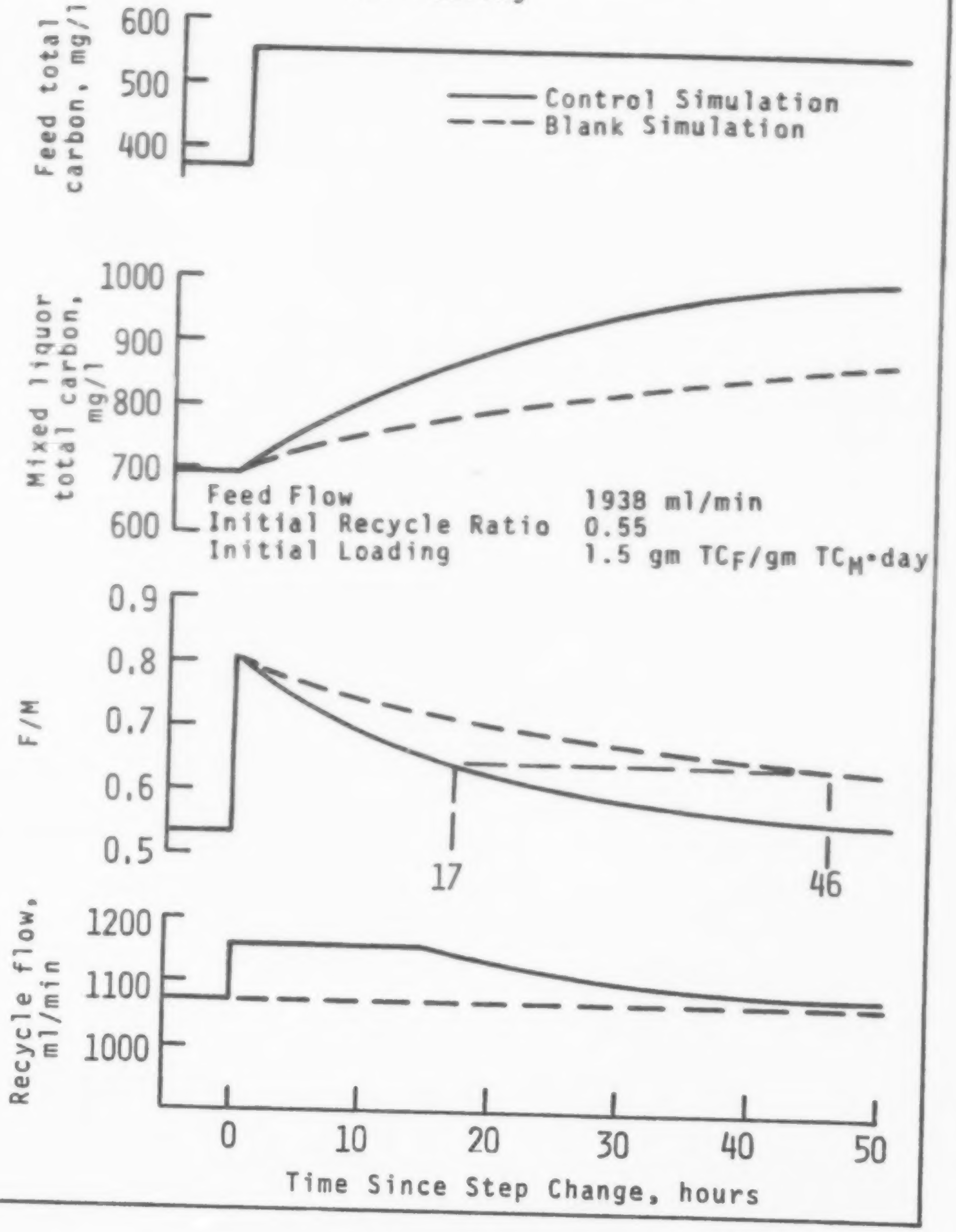
Figure 30 shows the relationship between substrate concentration in the aeration basin for the blank and the F/M control simulation of Figure 29. The control enabled a fast return to the desired effluent quality.

Figure 31 shows the simulation of an F/M blank test and an F/M control test for the brine propylene glycol waste water system. Because only 40 ml/min of waste was available for additional recycle, the response time with F/M control was 33 hours. While this was 54 percent less than the 72 hours it took for the blank system, on-line F/M control for the salt system did not seem justifiable. Manual adjustment of the recycle and waste flows were adequate for loading control because of the long response times involved.

The implication from the testing of the fresh water and salt systems was that biota with faster growth characteristics (i.e., a high yield coefficient) necessitated relatively high waste rates to remove the cellular production. Because the F/M response time was limited only by the amount of waste available for additional recycle, F/M loadings in these high yield systems can be well controlled. For example, a domestic waste was found (Benedek and Horvath, 1967) to have a cell yield coefficient of 0.67 mg VSS per mg COD removed (k_d - 0.07 day^{-1}). This compares with 0.21 mg VSS/mg TOD_{rem} for the fresh water-propylene glycol system (k_d = 0.06 day^{-1}). The specific rate of sludge growth reported by Benedek and Horvath was nearly three times that of the glycol system. The F/M ratio response time for a controlled system such as this would only be a few hours.

The problems associated with F/M control where the microbial population had slow growth characteristics were readily overcome. Use of an aerated sludge storage or stabilization tank supplied additional recycle in periods of high loading. Innovative schemes similar to those used in contact stabilization could be designed for optimum loading control. In existing facilities, biota from aerobic sludge digesters might be used for F/M control purposes.

Figure 28. Response characteristics with and without F/M control---low loading



Feed Flow 1938 ml/min
Initial Recycle Ratio 0.55
Initial Loading 1.5 gm TC_F/gm TC_M•day

Figure 29. Response characteristics with and without F/M control---high loading

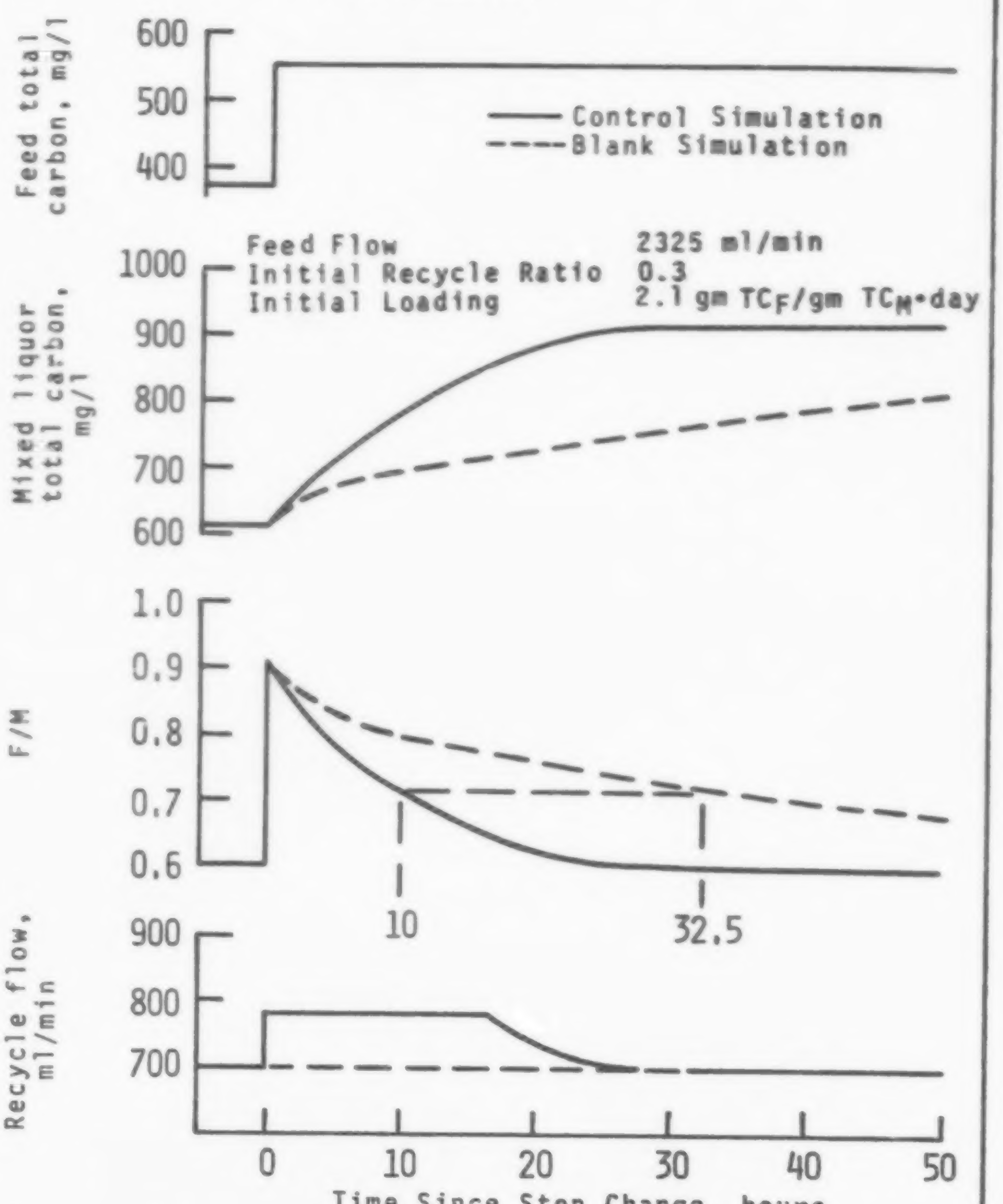


Figure 30. Comparison of aeration basin substrate concentration with and without F/M control

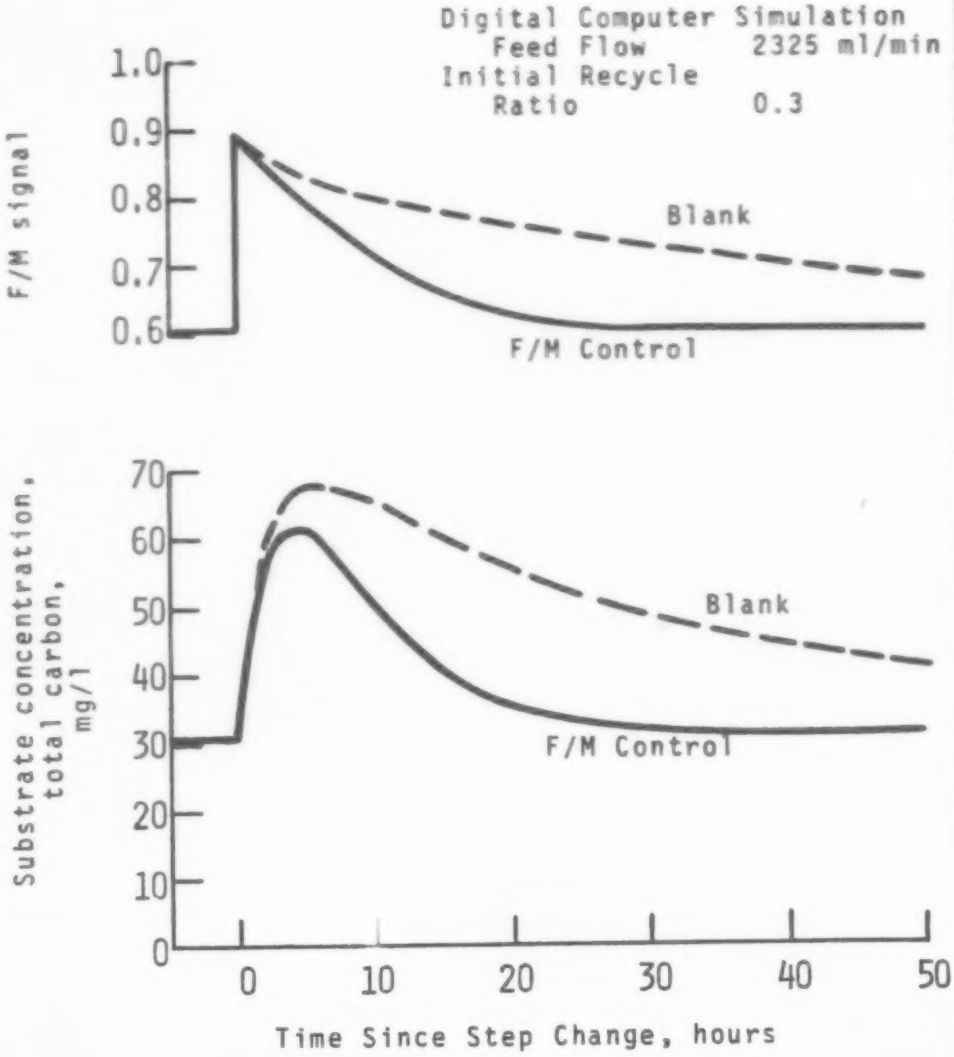
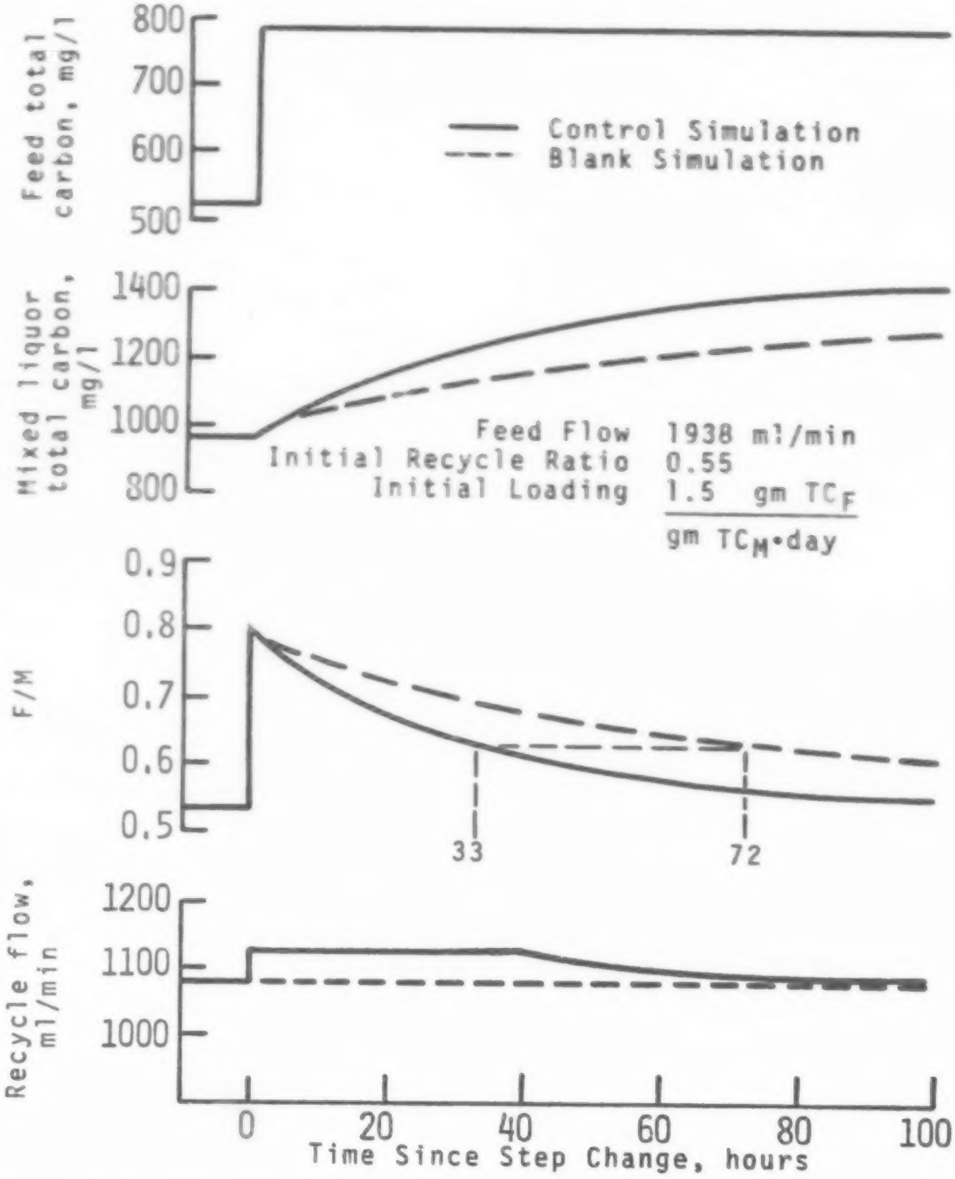


Figure 31. Response characteristics with and without F/M control (salt-propylene glycol system)



Feed Flow 1938 ml/min
Initial Recycle Ratio 0.55
Initial Loading $\frac{1.5 \text{ gm } TC_F}{\text{gm } TC_M \cdot \text{day}}$

SECTION VI

NUTRIENTS ADDITION CONTROL SYSTEM

An industrial waste stream to be treated by biological oxidation may not have sufficient concentrations of nitrogen and phosphorus to supply the amounts necessary for cell production and respiration requirements.

Constant nutrients addition, based only on flow, will occasionally result in excess or insufficient nutrients addition. Insufficient addition decreases bacterial viability while excess nutrients addition, besides being economically unsound, can cause downstream algae blooms and subsequent pollution problems. High nutrient residual may also prevent recycle of industrial treated effluents.

The solution to all of these problems lies in providing a varying flow of nutrients to the aeration basin based upon a measurement of the soluble organic substrate concentration in the feed.

SYSTEM DESIGN

The designed nutrients addition control system was operated on the basis of the addition of an ammonia and a phosphoric acid solution proportional to an analysis of total carbon in the feed stream. The feed sample was passed through a 3-micron Cuno filter and the total carbon content was measured by an Ionics Model 1212 Total Carbon Analyzer (TCA) operating on a six-minute analysis cycle. The recorder peak height was converted to an equivalent electrical signal of sufficient magnitude to match a peak picker sensor. The output of this sensor was transferred to a sample and hold amplifier by a programmer synchronized with the analyzer. The electrical signal was converted to a pneumatic signal which linearly adjusted the percent energized time of a 30-second cycle timer that powered a nutrients addition pump (see Figures 10 through 13).

The nutrients pump maintained a given flow until its controls received an up-dated analysis signal. If the period between analyses extended to hours or days, some slow reduction in the pump rate was anticipated due to decay in the sample and hold amplifier output.

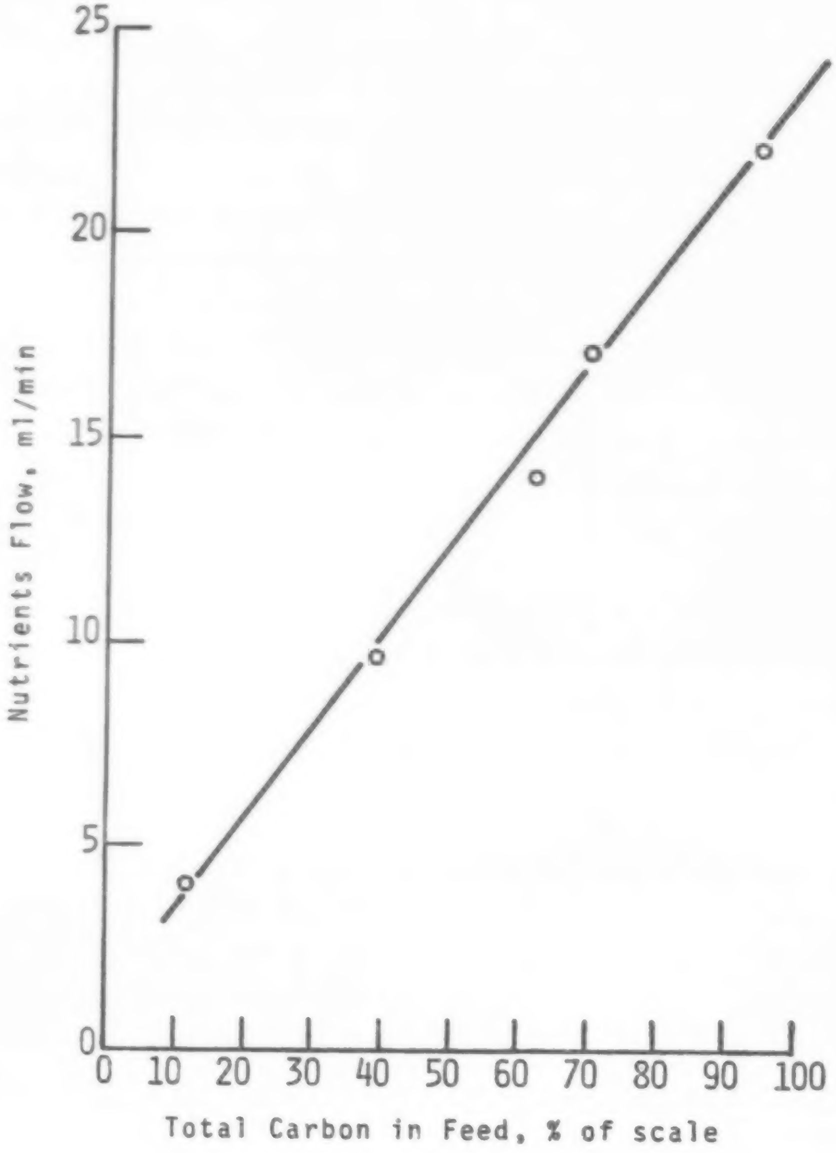
Proportionality of nutrients addition was varied by changing the influent flow, the carbon analyzer calibration, the full scale pumping rate, and the concentration of the nutrients solution. If required, additional variation in proportionality was available by altering the calibration of various intermediate hardware components.

The same total carbon analysis (TCA) that was used for F/M control (Section V) was also used for the nutrients control system. A flow chart and block diagram which includes both the F/M and the nutrients control system are given in Figure 10 and Figure 13, respectively. An electrical schematic of the controls is given in Figure 12.

SYSTEM TESTING AND PERFORMANCE

The time lag from the peak on the TCA to the air signal operating the nutrient pump was less than 10 seconds and the resulting linear response is shown in Figure 32. The control system performed reliably during the nine months of miniplant operation.

Figure 32. Response of nutrients control system



Routine analyses for the ammonia and phosphorus concentrations of the bio-settler overflow were made in order to determine the best C:N:P ratio for the miniplant operation. Ammonia and ammonium ion concentrations were determined by the Kjeldahl Nitrogen method. Total phosphorus concentration was determined by a molybdate colorimetric method, after co-precipitation of the phosphorus with iron.

Figure 33 shows the relationship between ammonia consumption and microbial growth. Data were calculated by material balance on a daily basis. Although the experimental data reflects the ammonia consumption for cellular growth plus the ammonia consumed by adsorption onto the bio-floc, the results compare closely with the theoretical ammonia consumption for cellular grwoth only. For a cell formula of $C_5H_7O_2N$, 0.15 grams of ammonia would be required for the growth of one gram of new cells. This compares with the experimental value of 0.133 gm NH_3/gm cells formed.

Figure 34 shows that the phosphorus consumption is directly proportional to the ammonia consumption. The phosphorus required is 0.06 gram per gram of ammonia. This translates to 0.008 gram of phosphorus required per gram of microbial cells formed.

The linear curves of Figures 33 and 34 illustrate the theoretical basis for feed-forward nutrients control. Because microorganism growth is directly proportional to the incoming feed concentration and because the nutrients consumption is directly proportional to the microorganism growth (Figure 33), the nutrient utilization will vary directly with the incoming feed concentration.

Figure 33. Ammonia consumption by microorganisms

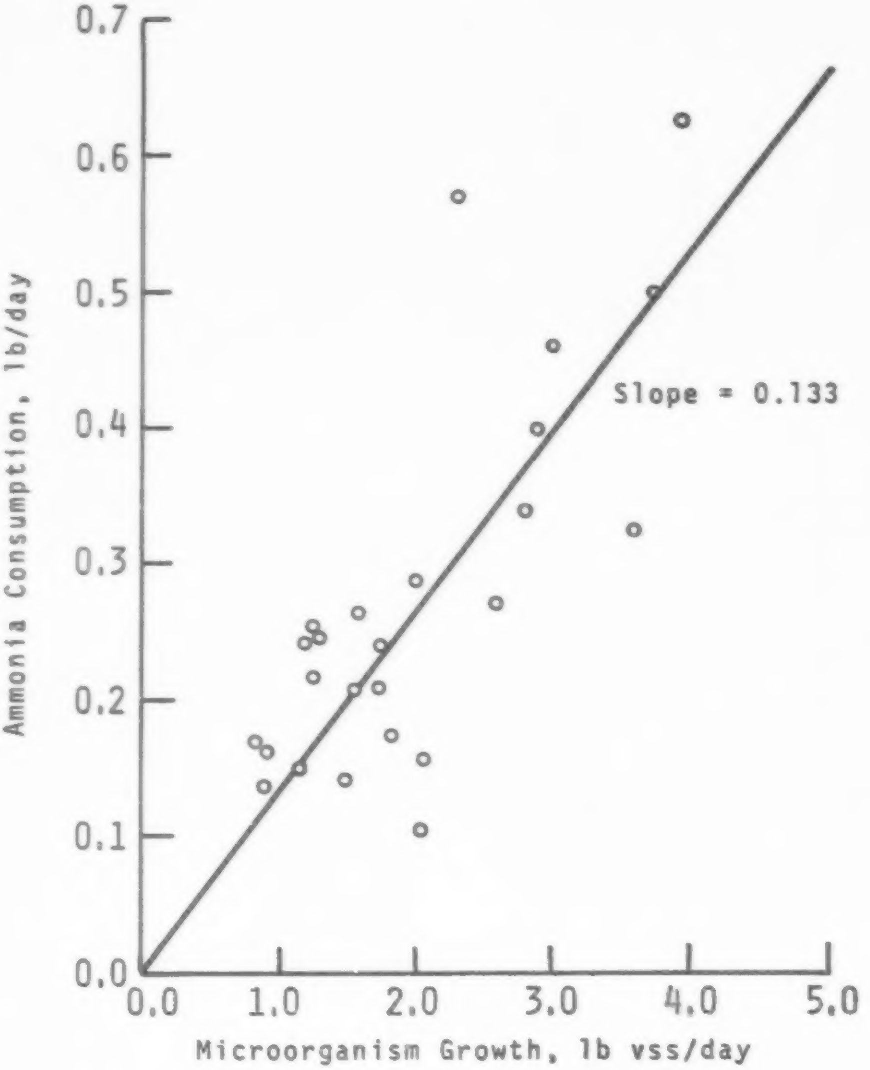
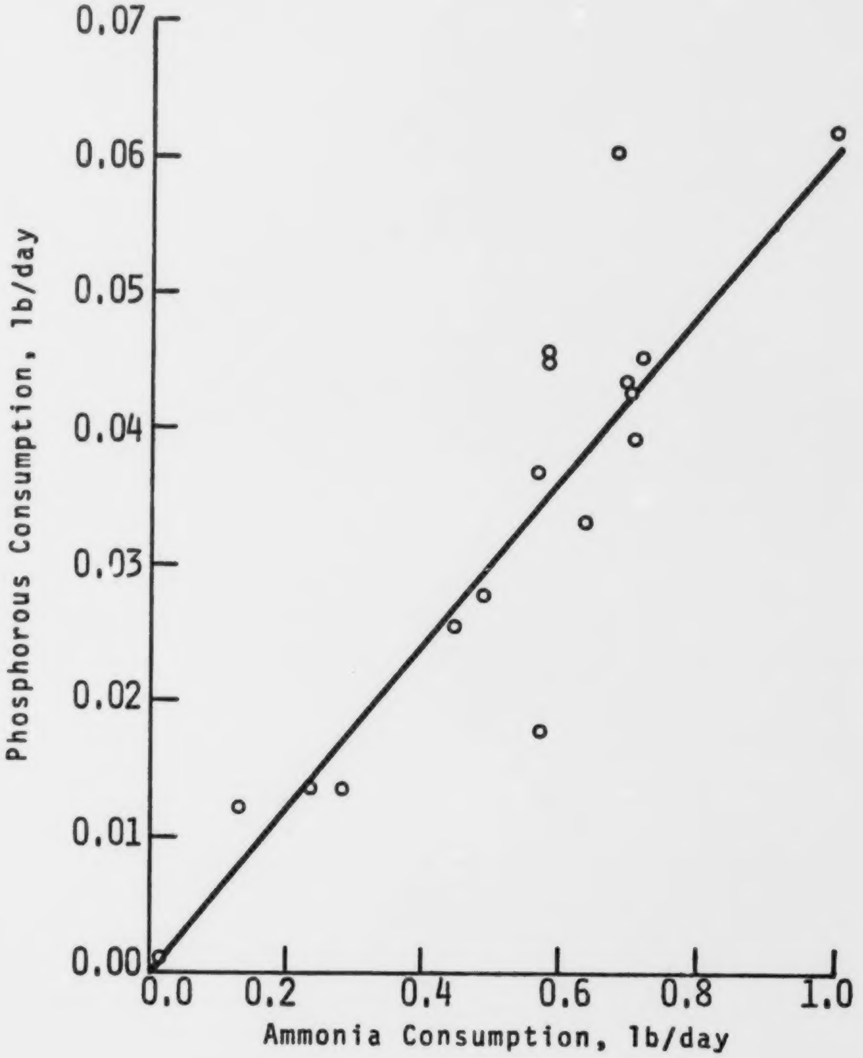


Figure 34. Relationship between phosphorous and ammonia consumptions

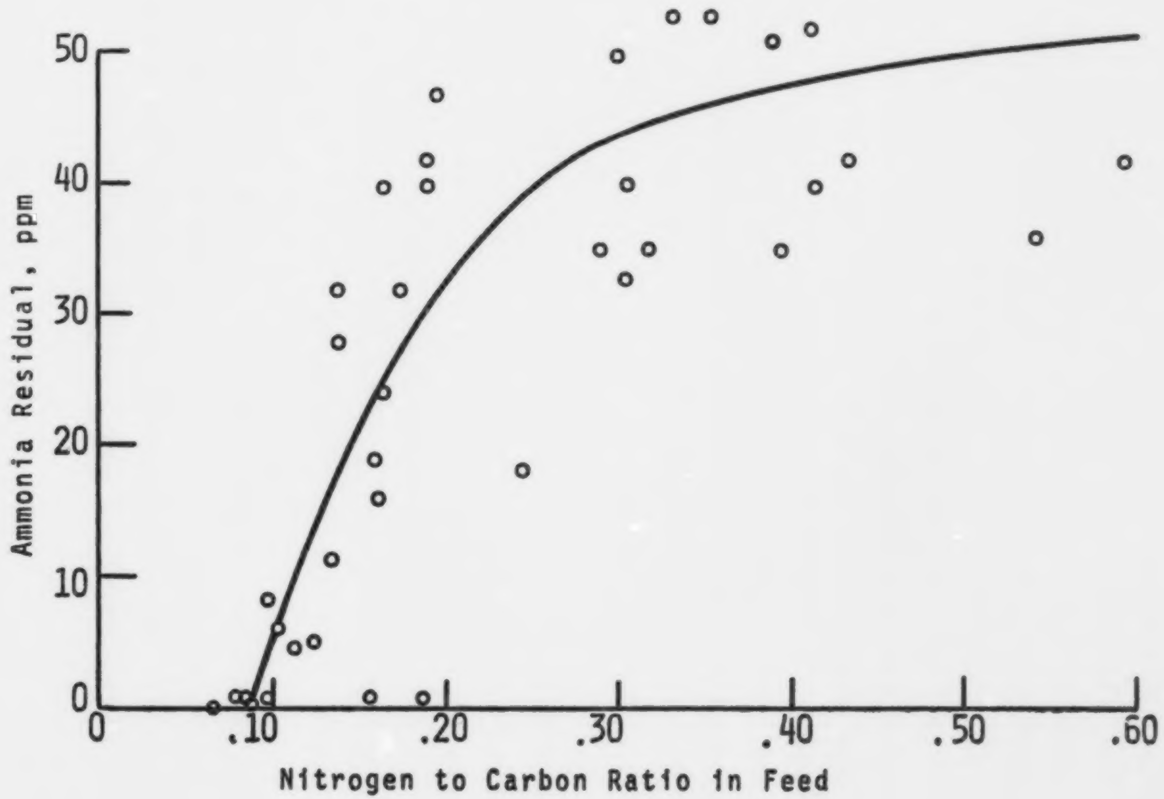


In order to determine the minimum nutrients requirement, the ratio of nutrients addition to feed total carbon concentration was progressively lowered until analysis of the bio-settler overflow indicated a low concentration of ammonia and phosphorus. The results shown in Figure 35 indicate a large adsorptive capacity of the mixed liquor bio-floc for values (nitrogen fed/carbon fed x 100) greater than 20. Material balance calculations show that without any bio-adsorption, for every ten units increase in the abscissa value above the minimum value, a 40 ppm increase in the bio-settler overflow ammonia concentration would have occurred. Because of the high adsorption, if the data of (N:C x 100) above 20 in Figure 35 were plotted in Figure 33 they would lie considerably higher than the linear curve presented.

The results in Figures 34 and 35 indicate an optimum carbon-to-nitrogen ratio of 100 to 9 and an optimum carbon-to-phosphorus ratio of 100 to 0.54. Using this proportionality the nutrients control system assured microbial viability and the effluent concentration was maintained at less than 10 ppm ammonia and less than 2 ppm phosphorus.

This control system could be modified to allow variations in influent flow rate by flow measurement and increased nutrients addition corresponding to the multiple of the flow times the feed total carbon analysis.

Figure 35. Determination of ammonia requirement



SECTION VII

CHEMICAL FLOCCULATION CONTROL SYSTEM

The overflow of the bio-settler is usually too turbid for release without some additional clarification. This is especially true in the treatment of saline waste waters, where overflow turbidities are unusually high.

In the pilot plant, a chemical flocculation control system was used to remove suspended and colloidal matter from both the fresh water and the saline biologically treated effluents.

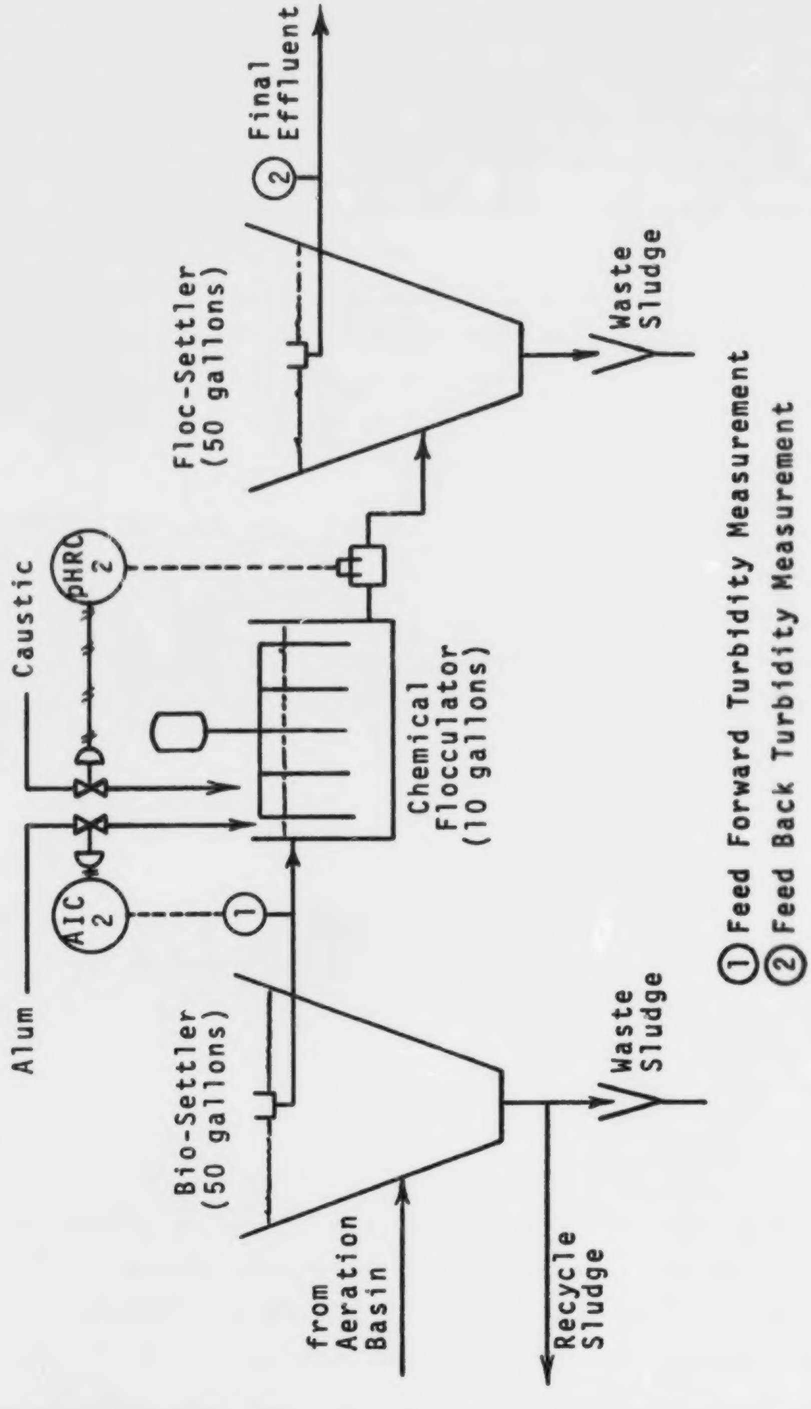
DESCRIPTION OF THE SYSTEM

The schematic in Figure 36 illustrates the flocculation controls and process flows. The proportional control was based on a turbidity measurement. Aluminum sulfate was used as the flocculant, and the pH was controlled (proportional-integral) at the level corresponding to the optimum for alum flocculation.

Both feed-forward and feed-back modes of control were tested. An Ecologic Model 204 Turbidimeter was used for feed-back control but fouling of the sensor window surfaces was a reoccurring problem. A timer-operated water flushing system was therefore installed to clean the probe every 100 minutes. This somewhat improved the instrument operation, but significant fouling still occurred.

The Ecological turbidimeter was replaced by a Hach Surface Scatter No. 2426 Turbidimeter in which there is no contact between the sample and the optical surface. In both feed-back and feed-forward control the entire settler overflow was passed through the instrument.

Figure 36. Flocculation controls


from Aeration Basin
Recycle Sludge
Bio-Settler (50 gallons)
Waste Sludge
Alum
AIC 2
①
Caustic
pHRC 2
Chemical Flocculator (10 gallons)
Floc-Settler (50 gallons)
Waste Sludge
②
Final Effluent
① Feed Forward Turbidity Measurement
② Feed Back Turbidity Measurement

The pH of the flocculator effluent was measured with conventional glass electrodes and recorded on a pneumatic recorder-controller. The caustic added was a 2-percent NaOH solution. A high pH alarm automatically closed the control valve at values above 8.3 and aided in control recovery under high pH-upset conditions.

The pilot plant chemical flocculation control system was operated at a constant overflow rate in the range of 1,800 to 2,200 ml/min.

OPTIMUM pH AND ALUM DOSAGE

The stability of bio-colloids is dependent upon the surface charge density. In general, the microorganisms carry a net negative charge, within the pH-range of interest. The charge is acquired through acid-base interactions of functional ionogenic groups and, thus, the biota surface charge density is strongly pH dependent (Tenney and Stumm, 1965).

Flocculation is not a result of the aluminum ion Al (III) but of its hydrolysis products (Stumm and Morgan, 1962, Stumm and O'Melia, 1968). The solution pH is the most important parameter in determining which particular polymeric hydrolysis specie predominates. This in turn influences the polymer bridging characteristics which cause flocculation.

Jar Tests

The optimum pH for alum flocculation was determined by a series of jar tests at various pH values. A one-liter sample of the bio-settler overflow was placed in a beaker and 1M HCl or NaOH was added to bring the sample to the desired pH. The pH was held constant by adding a predetermined amount of caustic simultaneous with the alum dose during a one-minute, rapid-mixing phase. After an additional slow stirring at 30 rpm for 15 minutes, the sample was allowed to settle for 20 minutes. The supernatant was then siphoned and analyzed for turbidity.

To determine the optimum pH for flocculation, the tests were performed on samples having the same initial turbidity. There was an increase in the optimum alum dosage for flocculation above pH of 6 and below pH of 5. Also, the height of the settled floc subsidence line was much higher when the pH was above 7 and when the alum doses were greater than 1 millimole per liter.

Optimum Alum Dose as a Function of Turbidity

The meaning of optimum alum dosage is illustrated in Figure 37. If the dosage of alum is less than the optimum, effluent quality will diminish, while doses above the optimum are economically unsound. Also, high alum doses were found to cause light, fluffy flocs, which sometimes resulted in settler upsets because of poor settling and high subsidence levels. This would result in less compacted sludge and necessitate increased sludge dewatering.

There exists a stoichiometric relationship between the bio-settler overflow turbidity and the optimum alum dose for flocculation, at constant pH (Figure 38).

The effect that the physiological condition of the bacteria has on the optimum alum dosage for flocculation is illustrated in Figure 38. Data point 5 represents a series of jar tests on a sample of bio-settler overflow with an artificially increased initial turbidity. The turbidity of the overflow

Figure 37. Alum flocculation jar tests

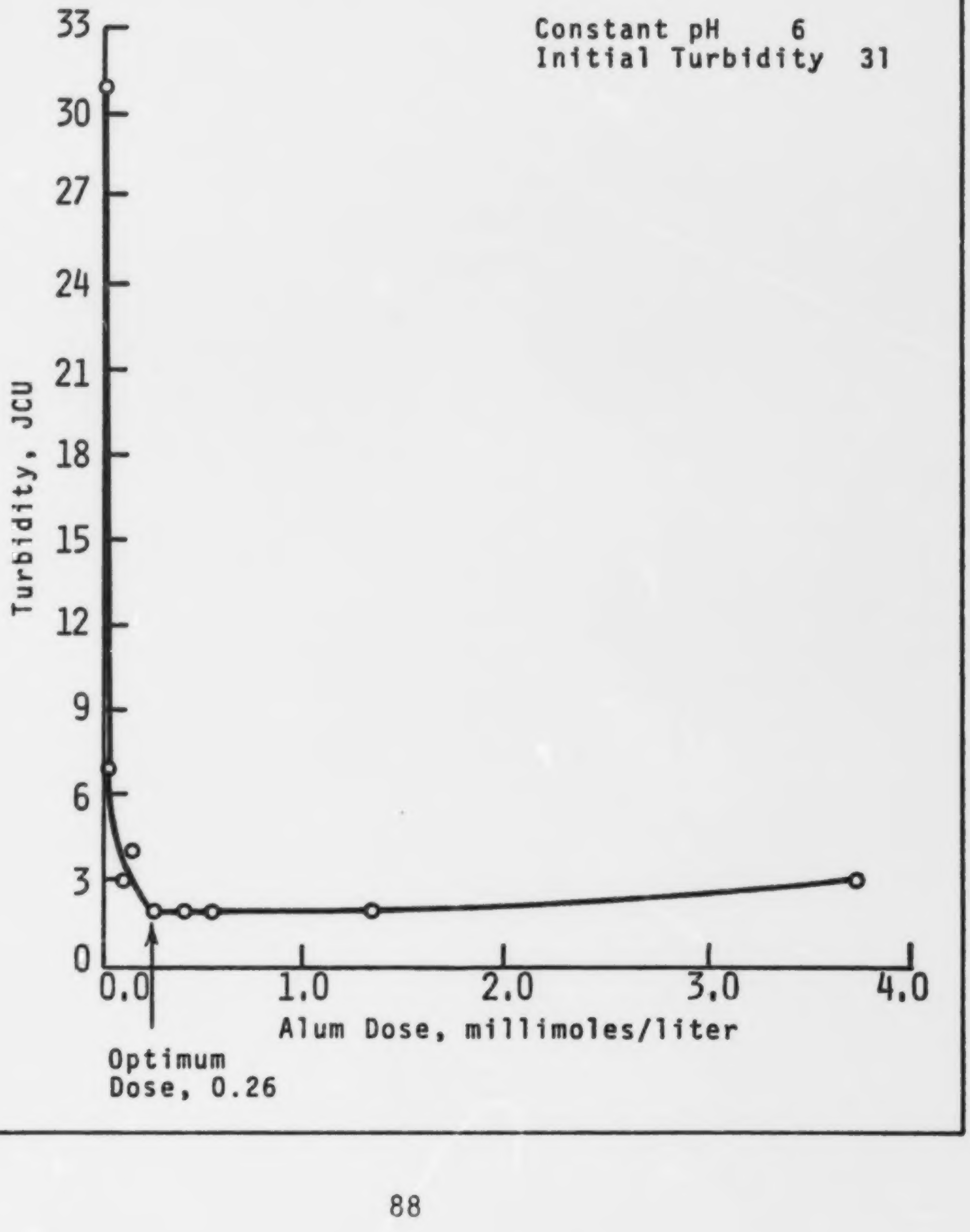
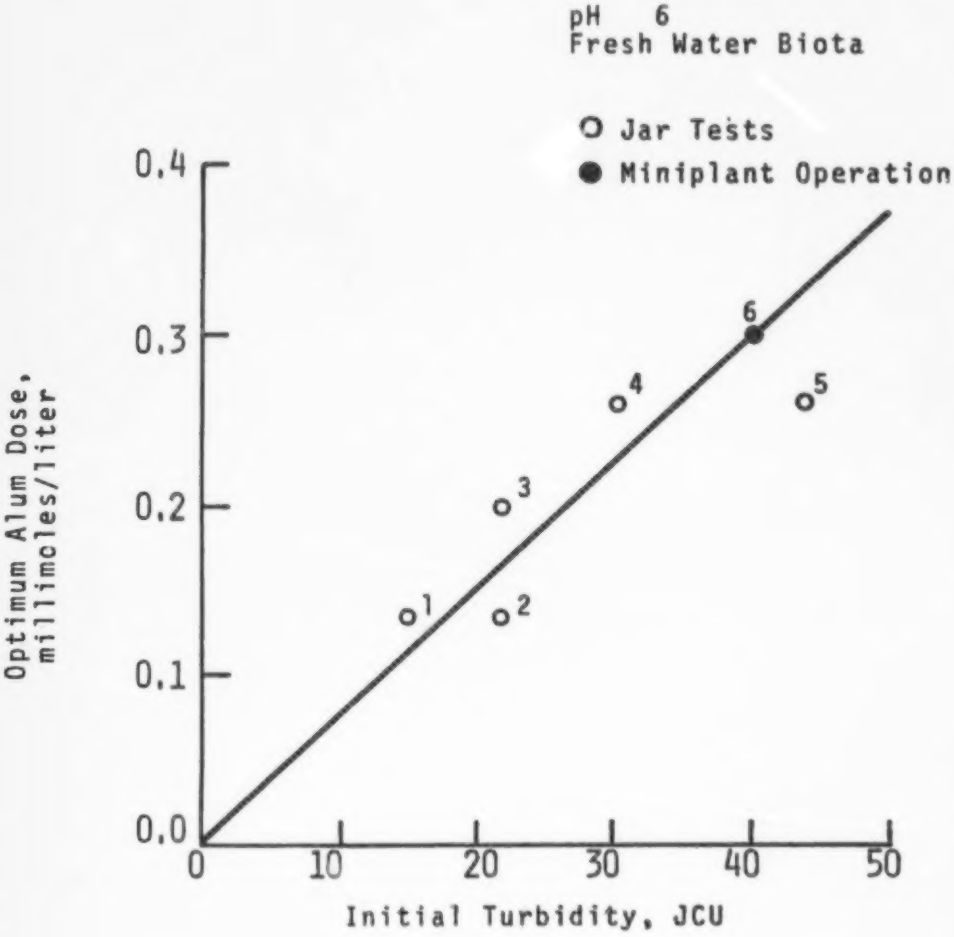


Figure 38. Optimum alum dose as a function of initial turbidity



sample was increased from 25 JCU to 44 JCU by adding aeration basin mixed liquor to the sample. Thus the sample represents larger bio-colloids of older microorganisms. On the other hand, data point 6, where a higher optimum alum dose was required, represents a case of finely dispersed microorganisms. Point 6 is taken from continuous flow miniplant data. Here the alum flow was increased just to the level of good flocculation. The high, bio-settler overflow turbidity was the result of a very high growth rate of microorganisms in the aeration basin.

SYSTEM OPERATION AND PERFORMANCE

The Hach Surface Scatter Turbidimeter worked well in both modes of control. When operating in the feed-forward mode it was necessary to drain the instrument every two to four hours in order to flush a buildup of bacteria on the inside walls of the turbidimeter body cylinder.

Because of the lag time of several hours associated with the flocculator and final settler, control stability was a major problem in feedback control.

Since in feedback control an increase in the floc-settler turbidity was necessary before the controller would increase the alum flow, final settler upsets were not uncommon. High alum additions because of a slight settler upset would sometimes result in a fluffy, poor-settling floc which would further upset the condition. High alum doses sometimes caused a milky white effluent, increasing the overflow turbidity and causing total instability.

The feed-forward mode of proportional control resulted in a high quality effluent. Stability was not a problem.

The feed-forward control was tested in both the propylene-glycol fresh water system and the glycol-salt waste water system. Typical performance of the control system is given in Table 4 and Table 5.

The calibration of the optimum alum dose versus overflow turbidity shown in Figure 38 was used for the feed-forward control for the fresh water system. A ten-percent higher calibration was necessary for the salt system.

Dosages of alum were sufficient to remove residual phosphate (about 2 ppm phosphorus) due to excess nutrient addition. Higher concentrations of phosphorus required additional alum in stoichiometric proportions.

An average of 98 percent of the suspended solids were removed by the flocculation step in the fresh water system and 85 percent were removed in the salt system.

The chemical flocculation control system increased process efficiency in terms of TOD removal. (TOD data was taken from samples passed through a 25-micron Cuno filter.) For the fresh water system, an average of 25 percent of the bio-settler TOD was removed in the flocculation step. This increased the overall process efficiency from 90.9-percent TOD removal to 93.4 percent. A 37-percent reduction in effluent TOD resulted in an increase in the average, overall process efficiency from 74.5 percent to 85.4 percent for the saline propylene glycol system.

The turbidity of the final fresh water effluent was reduced to 2 to 3 Jackson Candle Turbidity Units (JCU); whereas, the bio-settler overflow turbidity was typically 15 to 25 JCU.

Table 4. ALUM FLOCCULATION SYSTEM PERFORMANCE,[a]
GLYCOL-FRESH WATER SYSTEM

Day of Operation	Effluent Before Chemical Flocculation		Effluent After Chemical Flocculation	
	MLSS, mg/l	TOD Removal, %	MLSS, mg/l	TOD Removal, %
1	45	88.8	4	89.7
2	22	95.5	0	96.0
3	20	93.5	0	95.8
4	20	89.3	0	93.0
5	11	92.0	0	93.9
6	36	95.2	1	95.4
7	52	84.1	0	87.7
8	86	90.7	0	94.9
9	34	88.3	0	94.0
10	55	91.7	0	93.9

[a] proportional control with optimum alum dosage according to Figure 38

effluent turbidity maintained at 2 to 3 JCU

Table 5. ALUM FLOCCULATION SYSTEM PERFORMANCE,[a] GLYCOL-SALT WATER SYSTEM

Day of Operation	Effluent Before Chemical Flocculation MLSS, mg/l	 TOD Removal, %	Effluent After Chemical Flocculation MLSS, mg/l	 TOD Removal, %
1	91	76.2	24	91.9
2	164	78.9	17	91.9
3	52	63.2	0	65.9
4	43	66.6	3	----
5	23	90.3	0	90.8
6	214	75.8	18	96.3
7	66	71.5	0	----
8	43	69.0	19	----
9	52	75.4	21	80.3
10	40	78.4	18	81.0

[a] alum dosage at optimum (10% greater than doses in Figure 38)

The bio-settler overflow turbidity for the saline system was generally 35 to 80 JCU and the turbidity of the final effluent was lowered in the same proportion as the suspended solids removal, about 85 percent.

This control system could be modified to allow variations in the flow rate by measuring the flow and increasing the alum addition by the product of the flow times the turbidity.

SECTION VIII

BIOLOGICAL INHIBITOR DETECTOR

The kinetic reactions within a biological oxidation system proceed at rates defined by the nature of the chemical constituents in the waste water and the activity of the biomass. A real-time measure of enzyme activity to detect the presence of inhibitors or toxins in the feed to an activated sludge plant is required for optimization of the process.

The main objective of this study was to develop an upstream sensing device for toxic loads to an activated sludge process. Such a feed-forward control system is essential in an industrial or combined plants to prevent the loss or inhibition of the acclimated biomass. If the presence of a toxin in the feed to the plant can be detected rapidly enough on a repetitive basis, the feed could be diverted to a holding pond until the toxin is identified and eliminated.

BACKGROUND INFORMATION

Microbiological activity is a dynamic concept that expresses the ability of the microorganisms to interact with their environment. Inhibition of the activity of an acclimated culture may result from an adverse effect on the enzyme catalytic ability, the cell membrane permeability, or a general systemic effect by destroying the cellular integrity. Toxic, heavy metal ions affect enzymes by binding reversibly or irreversibly at the enzyme-active site or by causing conformational changes in the enzyme. Some organic inhibitors alter the cell membrane permeability, preventing the organisms from maintaining an environment favorable for proper metabolic activity. Heat and chlorine have a general systemic effect through denaturation of protein and other cellular components.

The enzyme theory of Michaelis and Menten (1913), used by biochemists to adequately describe the toxic effects on purified enzymes, was applied to describe the inhibition of activated sludge (Hartmann and Laubenberger, 1968). The application of enzyme inhibition equations to a mixed culture represents the overall kinetics of the biomass since it is not known if the toxicity effects are identical or different for various species of organisms. An inhibitor at a certain concentration may be toxic to some species but may stimulate the growth of others. The varying composition of the feed to an activated sludge plant adds to the problem in the study of toxicity. Some components may be synergistic or antagonistic effects with the toxin being studied (Poon and Bhayani, 1971).

The determination of different types of inhibitions of enzyme-catalyzed reactions with more than one substrate and product involves kinetic equations that are considerably more complex than the basic Michaelis-Menten equation (Cleland, 1963). Three types of inhibition patterns are commonly recognized from the kinetic analyses of experimental data:

1. Competitive inhibition results when inhibitor and substrate compete directly for the same enzyme site. Thus increased substrate concentration would reduce the effect of the inhibitor.
2. Uncompetitive inhibition is reaction of the inhibitor with the enzyme which can still undergo another reaction with the substrate. This type results in decrease of reaction velocity with increased inhibitor concentration since there is no direct competition between inhibitor and substrate for the free enzymes.

3. A mixed or noncompetitive inhibition is a mixture of the two mechanisms and is what takes place in a mixed culture such as the activated sludge.

The biological effects of toxic wastes have been conventionally evaluated by standard sanitary tests that measure the unit operation efficiency (BOD and COD reduction), biological population density (MLVSS and SVI), and biological oxidation capacities (sludge age, sludge yield, and rate of BOD exertion).

These measurements have been used by many investigators to assess the effects of toxic substances on the gross unit efficiency of an activated sludge system (Barth, et al., 1965; Banerji, et al., 1968; Ghosh and Zugger, 1973; Salatta, et al., 1964; Busch and Kalinske, 1956; Patterson, et al., 1969; Stack, 1956; Hermann, 1959; Smith, 1953).

None of these tests are appropriate for monitoring the activity of biomass in a treatment unit because of the time required to conduct any of these tests and the slow response of parameters to a reduction in the activity of the biomass caused by a toxic slug introduced into the treatment unit.

Biological waste treatment processes can be monitored through selective biochemical tests. Cells may exist in either of two metabolic states, the presence of utilizable substrate or in the absence of substrate. The endogenous metabolic level is indicative of the microbial population while the metabolic activity in the presence of substrate represents the oxidative activity of the biomass.

The microbial concentration if deoxyribonucleic acid (DNA) has been shown to be invariant with the physiological state and fairly constant among bacterial species (Genetelli, 1967). The concentration of DNA does not indicate the activity of a biomass and thus cannot be used as a measurement of toxic effects.

The hydrolysis of adenosine triphosphate (ATP), common to all cellular metabolism, is the basis of a method of measuring the physiological state of microbial cultures. The endogenous ATP concentration represents a relative measure of biomass while the rate of increase, or peak ATP concentration after substrate addition, reflects the biomass activity (Forrest, 1965 and Patterson, et al., 1969). Assay procedures for ATP are based on its bioluminescent reaction with luciferin and luciferase enzyme extracted from firefly lanterns. Briefly, ATP analysis requires rapid killing of the cells and extraction of ATP into aqueous solution. Many investigators utilize conventional liquid scintillation counters for measurement of luminscence. Others have developed their own light-measuring devices, and one company (E. I. DuPont) is now marketing a special instrument for ATP analysis. Concentration of ATP seems an attractive parameter to study the response of cells to their environment. Few attempts have been made to apply it to heterogenous systems such as an activated sludge, and its application as a continuous monitor does not seem practical.

The dehydrogenase activity test uses an electron acceptor dye 2,3,5-Triphenyl tetrazolium chloride (TTC), which is reduced to an insoluble red precipitate, formazan. The drhydrogenase activity is reported as micrograms of formazan produced during the incubation period per mg of cell material (Lenhard, 1965). The results of this test are greatly affected by incubation time, temperature, mixing rate, and the presence of oxygen (Patterson, et al., 1969).

Measurement of dehydrogenase activity in sludges was used to assess the oxidative capacity of activated sludge units (Ford, et al., 1965). The measurement was found to respond to significant changes in plant loading and sludge age, although no correlation between BOD removal efficiency and average dehydrogenase activity was found.

Many of the disadvantages of the dehydrogenase activity method are obviated by the direct measurement of the oxygen uptake, a measurement of the natural terminal electron acceptor without interrupting the normal biochemical processes.

Traditionally, the Warburg respirometer has been used in both biochemical and sanitary engineering research to measure the oxygen uptake rate of microorganisms. The Warburg technique requires special equipment and skilled technicians and is inappropriate for routine plant monitoring. Other manometric respirometers have been developed for waste treatment application (Arthur, 1964; Tool, 1967). The Arthur respirometer recycles the waste culture and air counter-currently through a large, closed, reaction chamber. The air-stream is passed through a sodium hydroxide solution to absorb carbon dioxide. As oxygen is utilized by the culture, the partial pressure of oxygen decreases within the aeration chamber, which is monitored by an oil manometer, and a pressure transducer. The AC signal from the transformer is then rectified and recorded as the change in partial pressure in the aeration chamber withtime, which represents the oxygen demand curve.

Continuous manometric respirometers have been developed for BOD monitoring (Clark, 1961; Pipen, 1973). In these devices, the oxygen uptake of a large, biological sample is continuously replaced by a manometrically triggered electrolysis reaction. The amount of electrical energy required to generate the required oxygen is recorded with time and is a measurement of the oxygen demand of the sample.

Since the development of membrane electrode systems to measure dissolved oxygen, such devices have been used to measure the oxygen uptake rate replacing the manometric methods.

The Simcar Respirometer (Abson, 35 al., 1967) utilizes a galvanic cell oxygen electrode to obtain a continuous record of dissolved oxygen concentration in an aerated culture to obtain the oxygen uptake of large samples over a period of 10 to 18 hours.

An automated respirometer (Wallace and Tierman, 1968) measures the oxygen uptake in a biological sample with an inlet probe and an outlet probe as the sample is pumped from one to the other in a predetermined transit time. The differential output of the two probes is representative of the oxygen uptake rate of the sample flowing through the device. A plot of the oxygen uptake rate versus time is then produced and the area under the curve represents the BOD of the sample tested. Results could be obtained within four hours for each test.

Oxygen uptake rates, both Warburg and oxygen electrodes, have been used to measure the effect of toxins and inhibitors on activated sludge (Ayers, et al., 1965; Ghosh and Zugger, 1971; Sato, 1971; Hartmann and Laubenberger, 1968).

The oxygen utilization rate shows more relative inhibition than the ATP or the dehydrogenase activity measurements, and it better approximates the actual activated sludge process (Patterson, et al., 1969). The oxygen utilization rate response rapidly to induced disruption of metabolic activity. It is sensitive, reliable, and can be easily automated with oxygen electrodes. A quantitative evaluation of commercially available oxygen probes has been recently published (Pijanowski, 1971).

OXYGEN UPTAKE MEASUREMENT CYCLE

Laboratory tests were conducted to determine an oxygen uptake measurement cycle that would be sensitive to the presence of a toxic substance in the feed to an activated sludge plant and that could be automated to rapidly detect the toxic effect on a repetitive basis.

A New Brunswick Scientific Chemostat Model C-30 was used as a batch reactor in which the temperature, mixing speed, and rate of aeration were controlled. It was also equipped with an analyzer and recorder to follow the dissolved oxygen (DO) level in the liquid phase in the reactor. A procedure was followed to measure the oxygen uptake rate and total oxygen utilized when a sample of a degradable carbon source was added to an acclimated and stabilized bacteria sample obtained from the activated sludge pilot plant:

> About 500 ml of the activated sludge was introduced in the reaction vessel and aerated to reach an equilibrium DO level that represented the endogenous equilibrium oxygen concentration. A 10-ml sample of a standard propylene glycol solution or a test sample containing a toxic substance was introduced in the reaction vessel

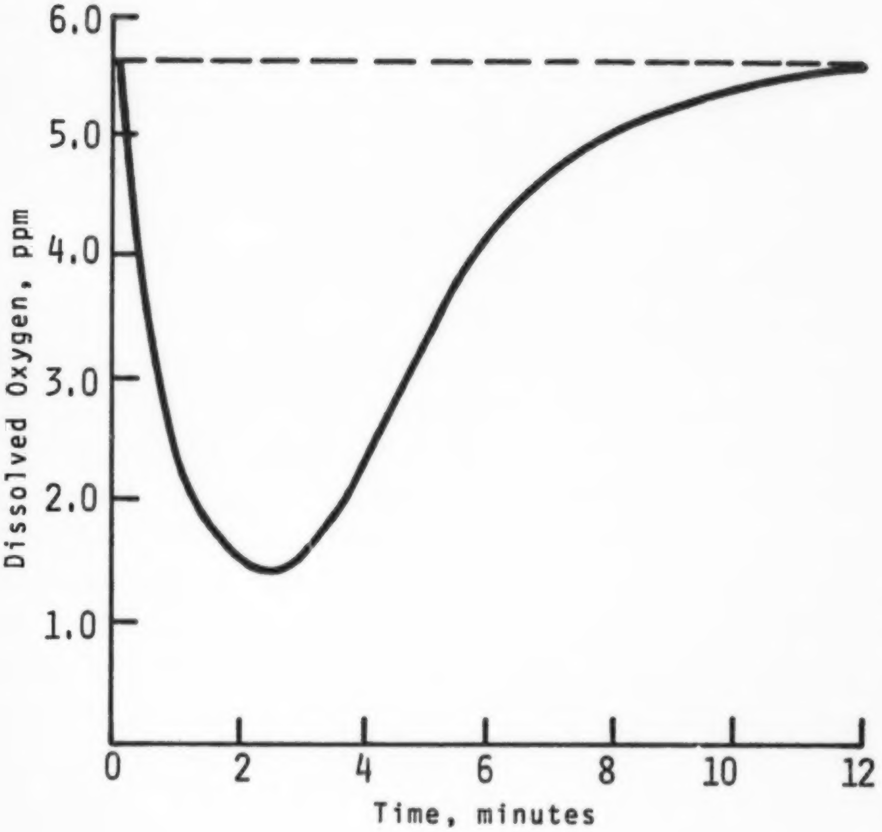
and the airflow turned off at the same moment. The initial slope of the decrease in DO concentration was the oxygen uptake rate due to the oxidation of the organic substrate plus the endogenous uptake rate. If a low airflow was maintained as the sample was introduced in the reactor, the oxygen concentration followed a curve as shown in Figure 39. The area under this curve was a measure of the oxygen consumed in the oxidation of the organic substrate in the sample added.

The volume of the test sample of 10 ml had negligible dilution effect on the DO concentration in the reaction vessel, and for a 1000-mg/l propylene glycol sample, complete degradation time was less than 15 minutes.

The oxygen uptake rate and the total oxygen consumed by propylene glycol standard solutions containing various concentrations of toxic substances were compared to the measurement obtained in the absence of the toxin. The sensitivity and reproducibility of the total oxygen consumed, as determined from the area under the oxygen uptake curve, was superior to that of the oxygen uptake rate, as determined from the initial slope of the oxygen concentration curve when the aeration was cut off. The rate of oxygen uptake was more dependent on the concentration of viable biomass than the total oxygen consumed. The automation of a measurement cycle utilizing the area under an oxygen uptake curve was simpler since it did not require cutting of the air every time a test sample was added.

For these reasons, the total oxygen consumed by a test sample added to an aerated sludge under controlled conditions was selected as the parameter to monitor toxic and inhibitory effects of the feed to an activated sludge process.

Figure 39. Schematic of oxygen concentration during the bio-oxidation of a degradable carbon source



Mathematical Model of the Oxygen Uptake Curve (Figure 39)

When a sample of a degradable carbon source was added to a stabilized and aerated sludge, the dissolved oxygen decreased rapidly due to the oxygen consumed by the bacteria to oxidize and metabolize the substrate added. The DO continued to decrease as long as the oxygen uptake was greater than the oxygen transferred to the solution by the controlled aeration. The time required to reach the minimum oxygen level depended upon the concentration and activity of the biomass, the organic loading, the rate of aeration, and the oxygen transfer characteristics. As most of the substrate was degraded and the rate of oxygen uptake became less than the oxygen supply, the DO started increasing till it leveled off at the original endogenous respiration equilibrium concentration.

The concentration of dissolved oxygen in the liquid phase, C, at any time is given by:

$$\frac{dC}{dt} = K_{La}\,(C_s - C) - r_r \qquad (16)$$

where K_{La} = mass transfer rate of oxygen, min^{-1}

C_s = saturation DO concentration, mg/l

r_r = oxygen uptake rate by the microorganisms, mg/l•min

Before the addition of the test sample containing the degradable substrate, the oxygen uptake by the microorganisms is the endogenous uptake and is proportional to the biomass concentration, X.

$$\frac{dC}{dt} = K_{La}\,(C_s - C) - K_eX \qquad (17)$$

At equilibrium $\frac{dC}{dt} = 0$ and therefore:

$$C_o = C_s - \frac{K_e X}{K_{La}} \qquad (18)$$

where K_e = endogenous reaction rate constant, min^{-1}
X = concentration of the biomass, mg/l

When a substrate is added to the biomass, the oxygen uptake will be the sum of the endogenous uptake and the uptake due to the substrate degradation. In a system that is not oxygen limited, the substrate utilization can be represented as a first order reaction by:

$$\frac{dS}{dt} = -k'XS \qquad (19)$$

where k' = substrate biodegradation rate constant, $\text{min}^{-1}\ (\text{mg/l})^{-1}$
S = substrate concentration in oxygen equivalent units, mg/l

Assuming the biomass concentration, X, to be constant during the degradation of a small sample of a substrate, Equation (19) can be integrated to:

$$S = S_o\ e^{-k'XS} \qquad (20)$$

The DO concentration after the addition of the substrate is given by:

$$\frac{dC}{dt} = K_{La}\ (C_s - C) - K_e X - k'XS \qquad (21)$$

Substituting Equations (18) and (20) in (21) and rearranging:

$$\frac{dC}{dt} + K_{La}C = K_{La}C_o - k'XS_o\,e^{-k'Xt} \qquad (22)$$

This differential equation is readily solved using an integrating factor, exp. $-\int K_{La}dt$ and the initial condition of Equation (18) to obtain the following expression for the DO concentration as a function of time:

$$C = C_o + \frac{k'XS_o}{k'X-K_{La}}\left(e^{-k'Xt} - e^{-K_{La}t}\right) \qquad (23)$$

The area of the oxygen uptake curve of Figure 39 is given by:

$$\text{Area} = \int_{t=0}^{t=\infty} (C_o - C)dt$$

$$= \frac{k'XS_o}{k'X-K_{La}} \int_{t=0}^{t=\infty} \left(e^{-k'Xt} - e^{-K_{La}t}\right)dt \qquad (24)$$

Performing this integration yields

$$\text{Area} = \frac{S_o}{K_{La}} \qquad (25)$$

The area of the oxygen uptake curve can be used to determine the biological oxygen demand (S_o) of the test sample if the oxygen mass transfer coefficient K_{La} is known. The value of K_{La} depends on the rate of aeration and mixing and the characteristics of the stabilized culture used in the test.

Development of the Measurement Cycle

In order to utilize the area under the oxygen uptake curve to monitor the toxic effects of a feed to an activated sludge plant, it is required to be able to distinguish between the changes due to the varying concentrations of the substrate in the feed and the presence of toxic substances. The oxygen uptake of a feed sample could be calculated per unit total organic carbon in the sample to eliminate the change due to varying substrate concentration, but this requires the use of another instrument to monitor the total organic carbon in the feed and a complicated, synchronized, electronic system.

A comparison of the oxygen uptakes of two standard propylene glycol samples before and after the exposure of the bacteria to a feed sample was a measure of the toxic effects of the feed to the activated sludge process. The oxygen uptake of the first standard established a reference activity of the bacteria sample, which was then exposed to the feed to the process for long enough time to degrade the substrate in the feed. Then the same bacteria sample received an equal volume of the standard and its oxygen uptake, compared to that of the first sample, was a meausre of any change of activity due to the feed sample.

The procedure developed for the automated biological inhibitor detector was as follows:

1. About 500 ml of the mixed liquor was withdrawn from the aeration basin to the reaction vessel of the instrument.

2. The bacteria sample was aerated at an adjusted rate of mixing and aeration for 16 minutes to reach an equilibrium endogenous respiration level.

3. A 10-ml sample of a 600 mg/l propylene glycol solution was introduced in the reaction vessel. During a period of 12 minutes, the oxygen uptake curve was recorded and its area, A_1, was determined and stored.

4. A 10-ml sample of the feed was then introduced to the reaction vessel and the DO level recorded over a period

of 18 minutes, enough to degrade feed sample containing substrates equivalent to about 1000 mg/l of propylene glycol.

5. Step 3 was repeated and the Area, A_2, of the oxygen uptake curve of this second standard solution was determined and stored.

6. The ratio of A_2 to A_1 was the activity ratio and was a measure of the toxic effects of the feed sample.

7. At the end of the cycle, the bacteria sample was drained from the reaction vessel and within two minutes a second cycle started. The total cycle time was 60 minutes.

Figure 40 is a schematic illustration of the measurement cycle when the feed contained no toxic substance. The activity ratio would be approximately equal to 1.0.

Figure 41 illustrates the measurement cycle when the feed contained a toxic substance. The activity ratio would be much less than 1.0.

LABORATORY TOXICITY TESTS

The developed method for the measurement of an activity ratio was used to study the effects of concentration and time of exposure of various toxic substances on a propylene glycol-acclimated fresh water bacteria. The purpose of the laboratory batch testing was to determine the response of the activity ratio measurement to the inhibitory effects of heavy metal ions and toxic organic substances that may be present in a petrochemical waste water treatment plant.

Figure 40. Biological inhibitor detector---schematic of measurement cycle with no toxin in feed

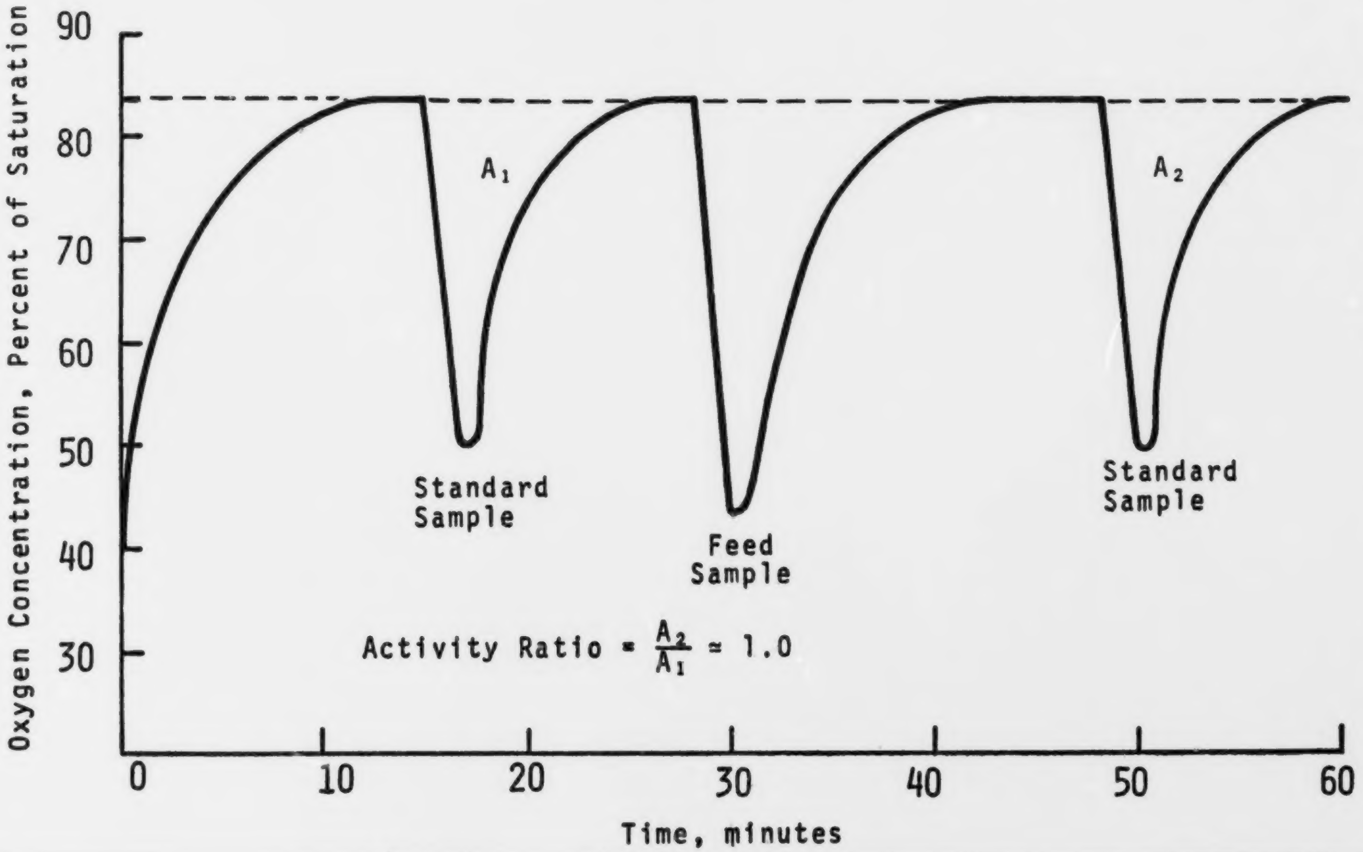
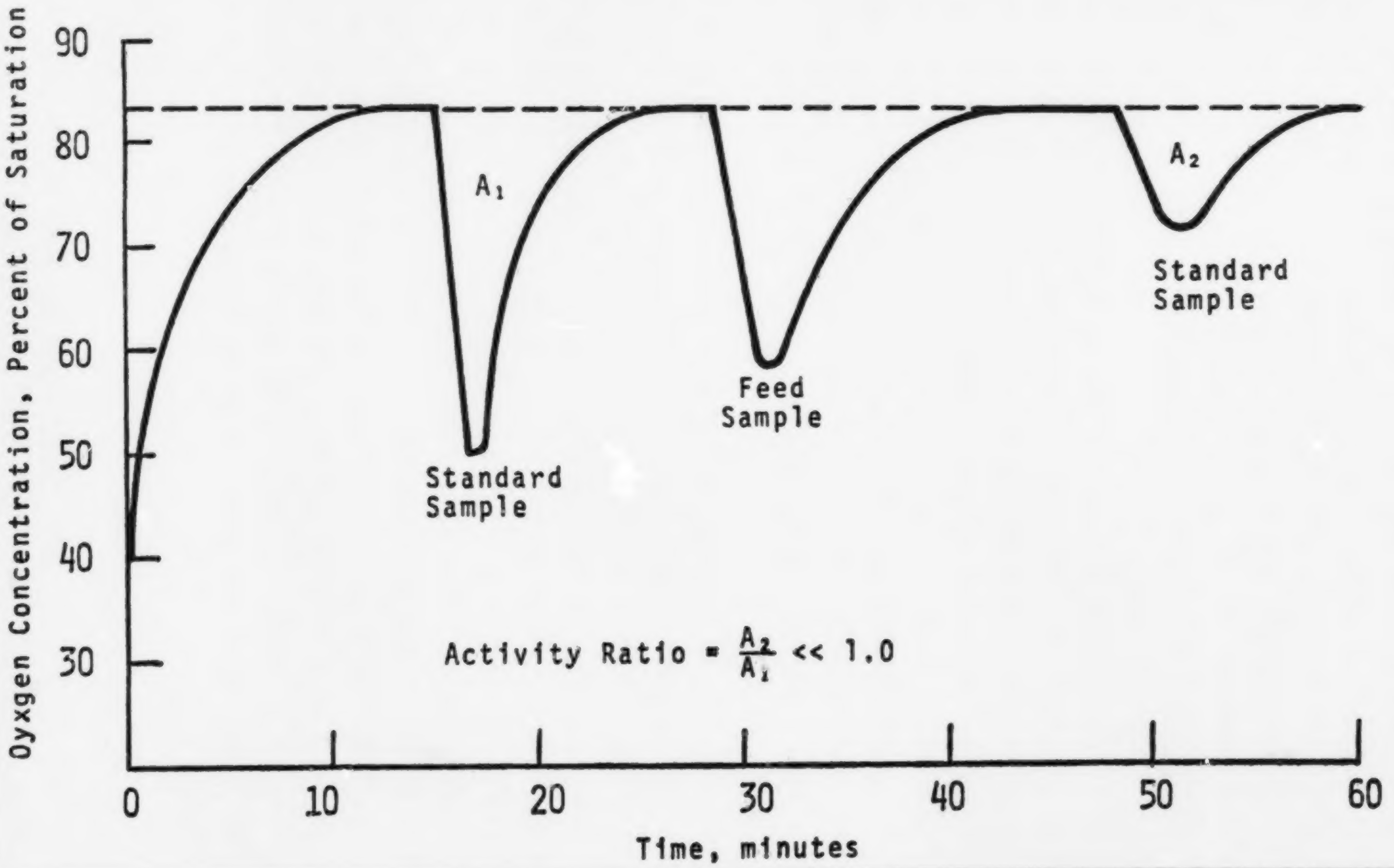


Figure 41. Biological inhibitor detector schematic of measurement cycle with toxin in feed



The bacterial culture, obtained from the activated sludge miniplant operating on a fresh water, propylene glycol feed, had a mixed liquor volatile suspended solids (MLVSS) of 1700 mg/l to 2800 mg/l during these tests. Direct plating of the acclimated culture indicated a microorganism composition of 77 percent *Pseudomonas*, 13 percent *Bacillus subtilis*, and 7 percent *Alcaligenes* as the main species.

The laboratory procedure was similar to the procedure developed for the monitoring instrument, in which the feed sample was a synthetic 600 mg/l propylene glycol containing varying concentrations of the toxic substance. The oxygen uptake of the second standard was measured at different time intervals after the exposure of the bacteria sample to the test solution to determine the accumulative toxic effects or if the bacteria recovers with time.

The oxygen uptake, calculated per unit MLVSS, for some selected inorganic toxins is plotted versus the concentration of the toxin in the test sample in Figure 42. The activity ratios for the same toxins are given in Table 6 for one concentration at varying exposure time. Mercury was the most toxic of the inorganic toxins tested, reducing the oxygen uptake to less than half at 3 ml/l and to nearly zero at 16 mg/l concentration. At an intermediate concentration of 8 mg/l, mercury exhibited an immediate reduction of the activity ratio to 0.27 that was further reduced with longer exposure times. The same chronic toxicity effect of the other inorganic toxins tested is illustrated in Table 6.

The oxygen uptake for samples containing varying concentrations of phenol and dichloroisopropyl ether is shown in Figure 43 and the effect of the exposure time on the activity ratio of the two substances is given in Table 7. Phenol was

Figure 42. Effects of inorganic toxins on the oxygen uptake of glycol-acclimated fresh water bacteria

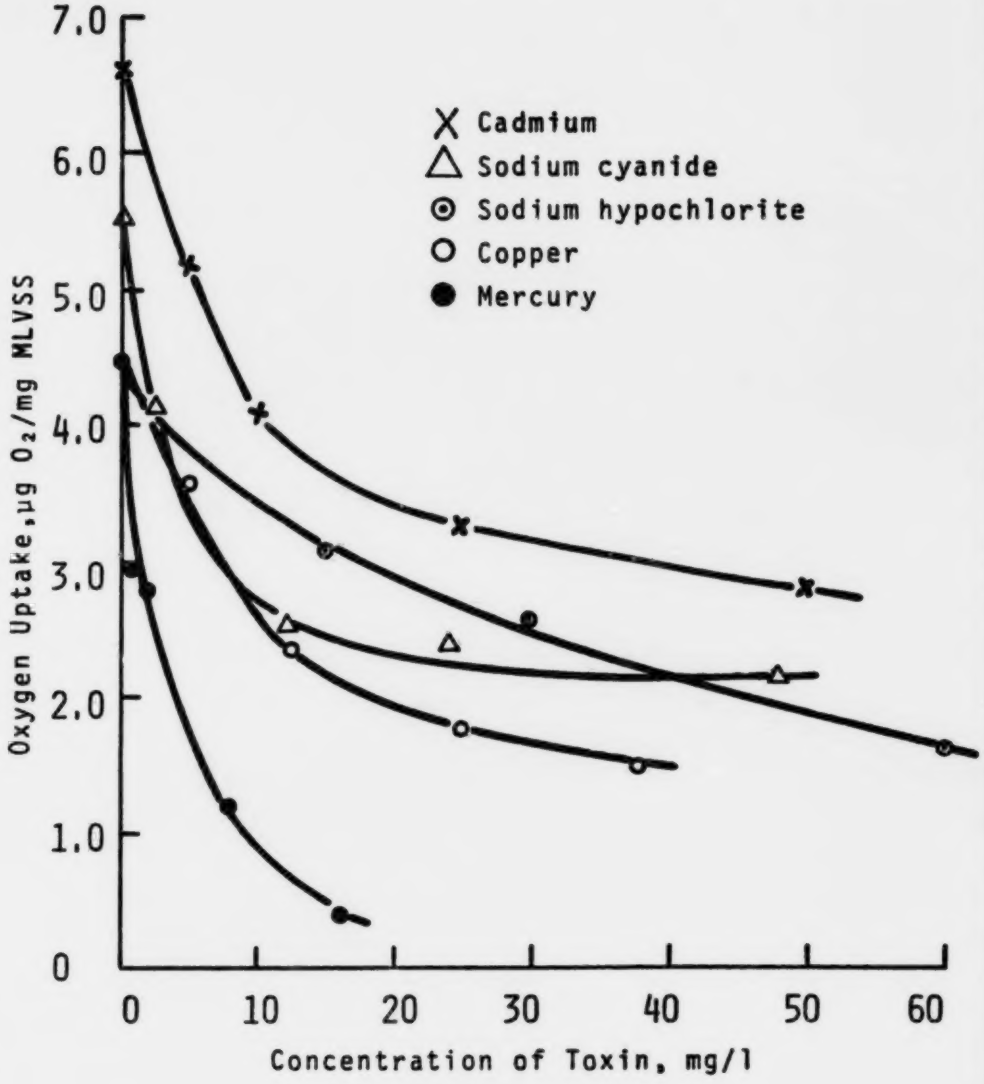


Table 6. EXPOSURE TIME EFFECT OF INORGANIC TOXINS ON GLYCOL-ACCLIMATED FRESH WATER BACTERIA

Toxin	Concentration mg/l	Exposure Time minutes	Activity Ratio (A_2/A_1)
Mercury	8	immediate	0.27
		16	0.20
		48	0.127
Copper	25	immediate	0.31
		19	0.206
		64	0.204
Sodium Cyanide	3.0	immediate	0.607
		20	0.502
		40	0.353
		50	0.316
Sodium Hypochlorite	60	immediate	0.294
		12	0.187
		25	0.102
Cadmium	50	immediate	0.70
		14	0.337
		24	0.276

Figure 43. Effects of organic toxins on the oxygen uptake of glycol-acclimated fresh water bacteria

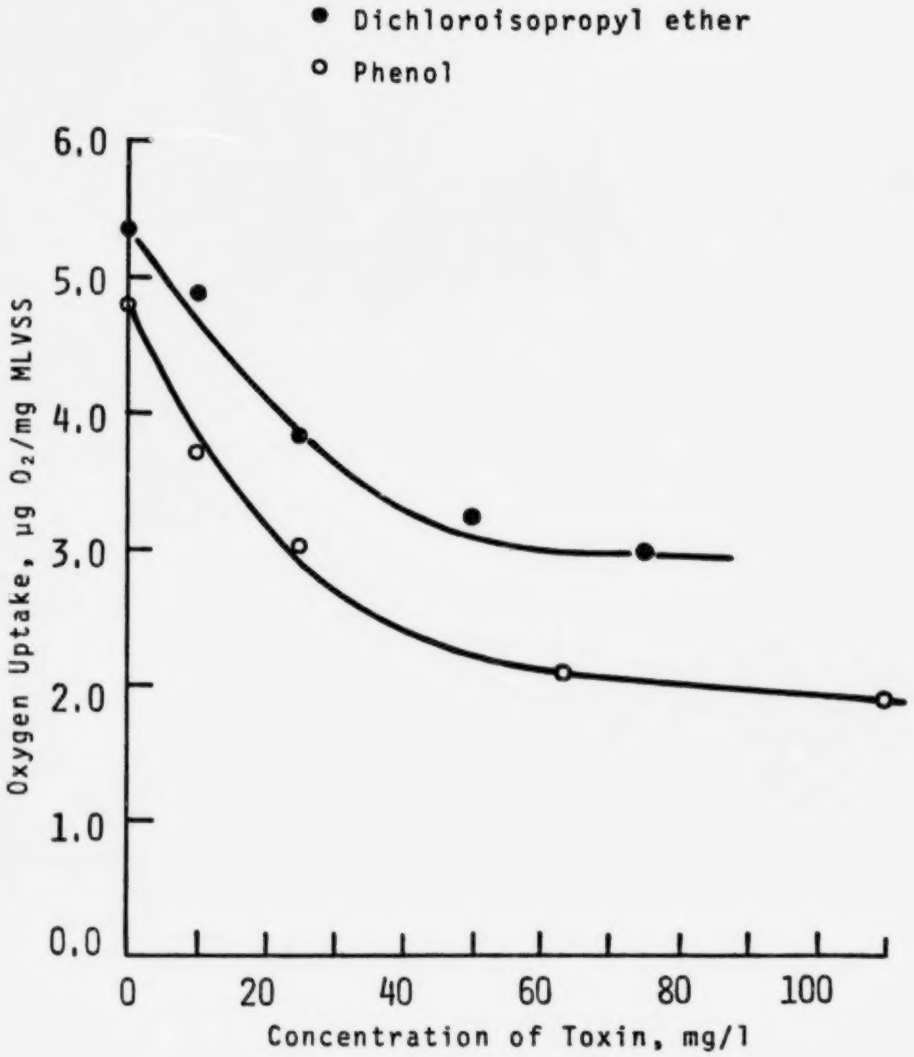


Table 7. EXPOSURE TIME EFFECT OF ORGANIC TOXINS ON GLYCOL-ACCLIMATED FRESH WATER BACTERIA

Toxin	Concentration mg/l	Exposure Time minutes	Activity Ratio (A_2/A_1)
Dichloroisopropyl Ether	100.0	immediate	0.498
		20	0.649
		40	0.649
		60	0.628
Phenol	111.5	immediate	0.394
		13	0.438
		27	0.417
		39	0.410

slightly more toxic to this culture than the ether, which occurred in low concentrations in the waste water from the production of propylene glycol. Also the recovery of the bacteria, as indicated by an increase of the activity ratio with time of exposure, was better for the ether than for the phenol. The immediate reduction of activity could have been due to a shock loading effect, followed by acclimated and, thus, recovery of the biomass.

DESIGN AND OPERATION OF THE AUTOMATED INSTRUMENT

The measurement cycle, developed and tested in the laboratory on standard solutions containing various toxic substances, was used as the basis for the design of the automated instrument.

The compact table top model, as built, is shown in the photographs of Figure 44 and Figure 45. The bacteria sample entered from the bottom connection to the reaction vessel and drained through the same connection. The reaction vessel was placed on a variable speed magnetic stirrer and an air sparger, sample lines, the oxygen probe, and a thermometer were introduced through the stopper. The air supply was controlled and measured by a rotameter while the exact volumes of the standard solution and feed samples were introduced through a calibrated, volumetric syphon system. A tracing of the measurement cycle as operated by a 10-cam timer is shown in Figure 46. A schematic of the electrical controls for the instrument is shown in Figure 47. A block diagram, Figure 48, traces the path of the measurement signal through the various electronic components. The recorder may be connected to any of six points indicated by the numbers in parenthesis in Figure 48 and listed in Table 8, the legend.

Figure 44. Biological inhibitor detector---electronics ard readouts

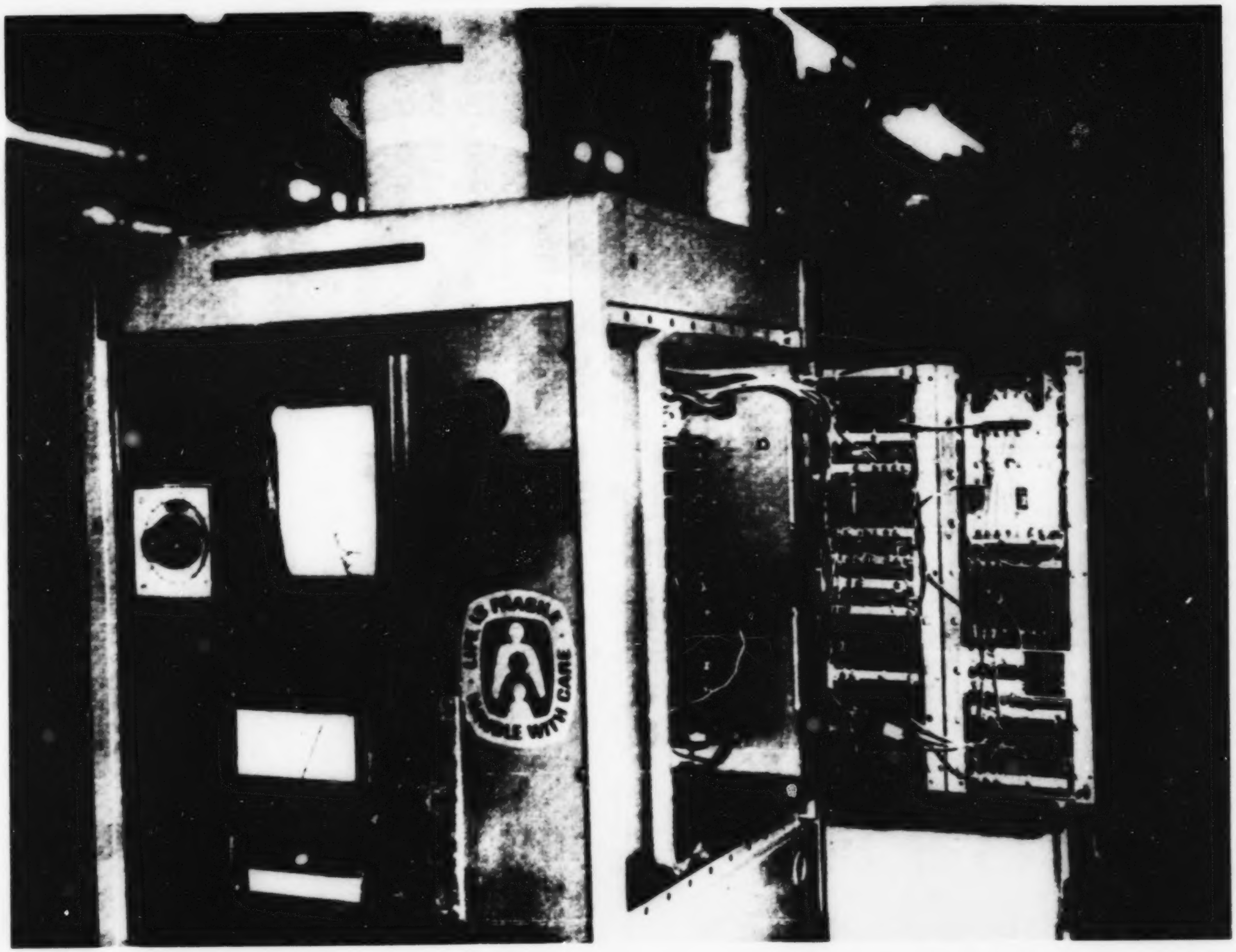


Figure 45. Biological inhibitor detector---reactor and sampling

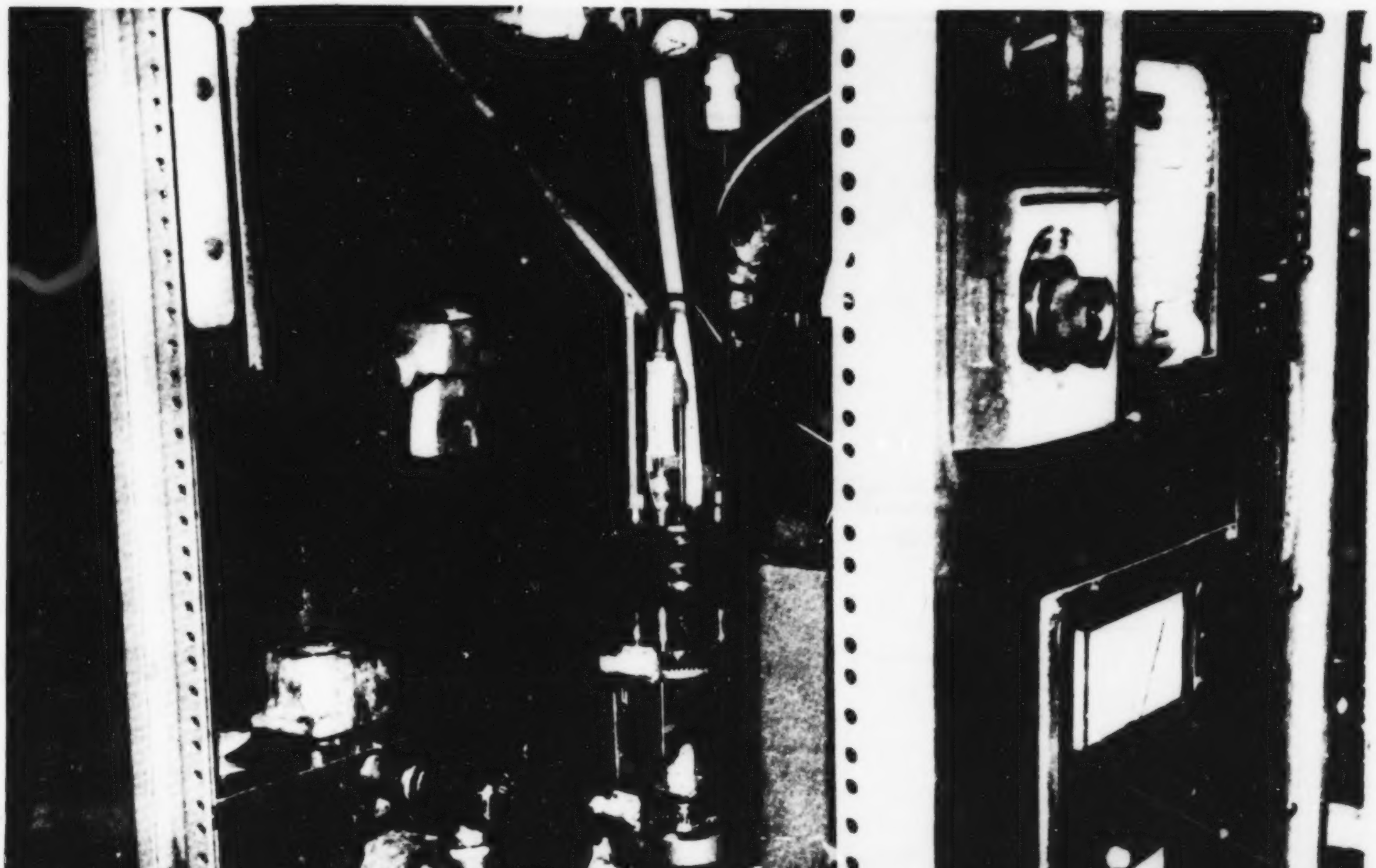


Figure 46. Biological inhibitor detector---cam time program

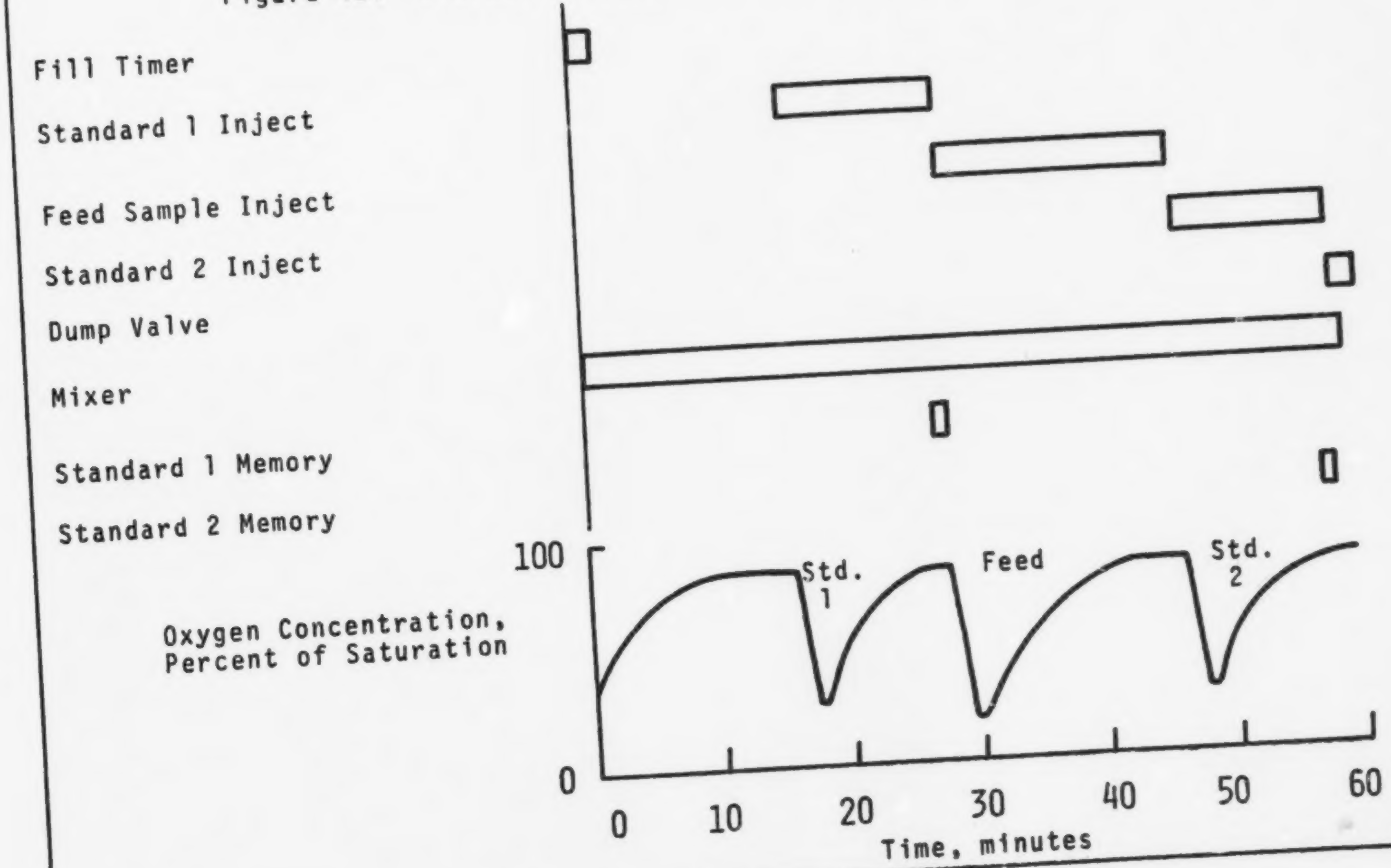
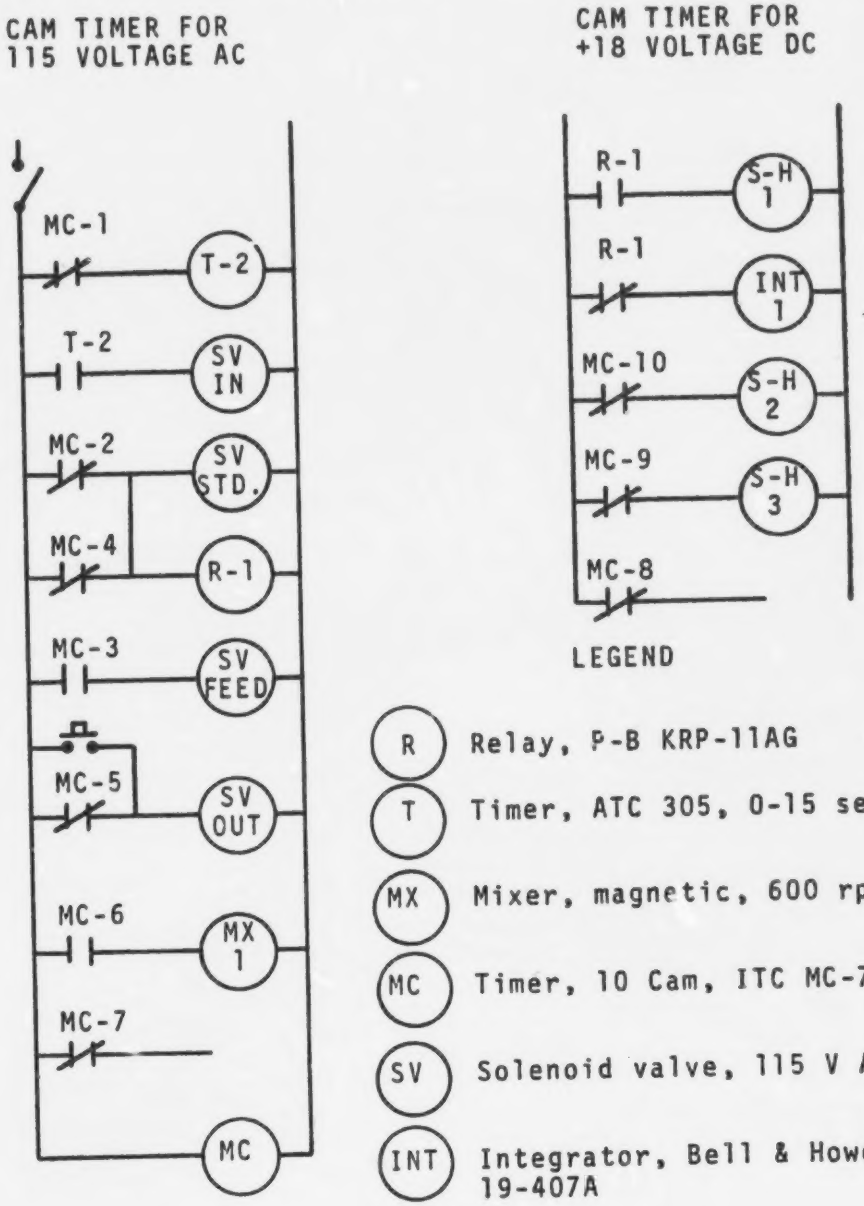


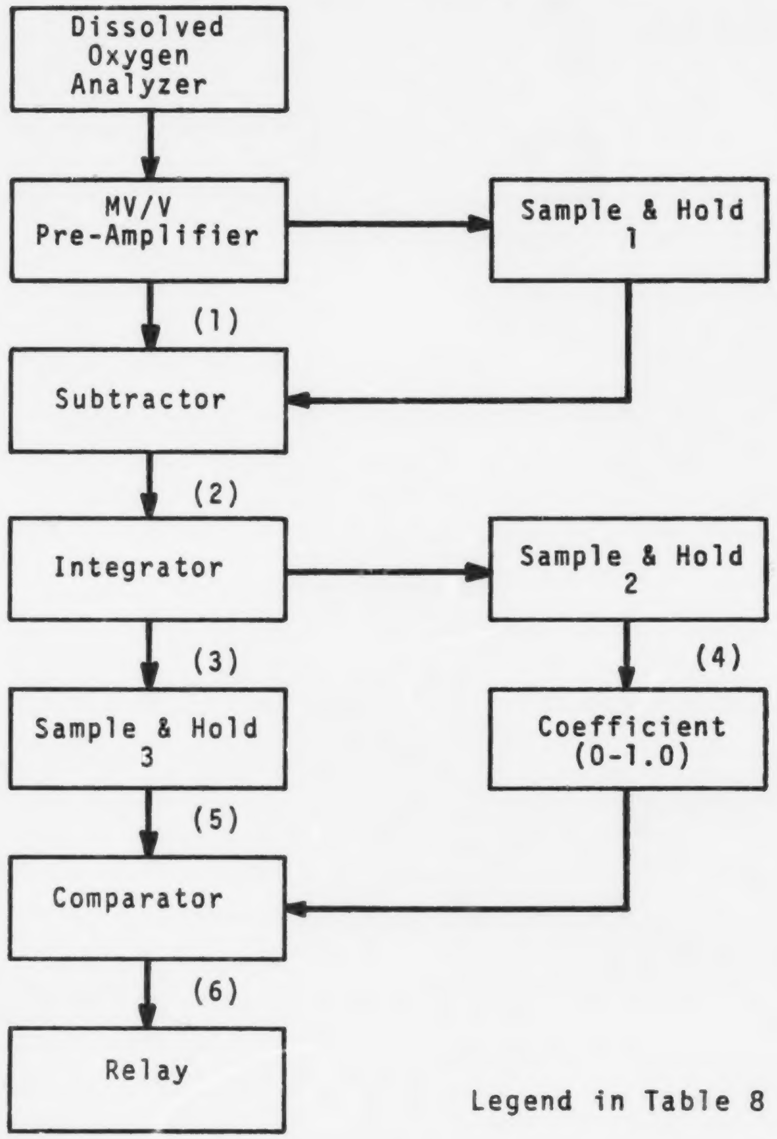
Figure 47. Biological inhibitor detector---schematic of electrical controls



LEGEND

R	Relay, P-B KRP-11AG
T	Timer, ATC 305, 0-15 seconds
MX	Mixer, magnetic, 600 rpm
MC	Timer, 10 Cam, ITC MC-7
SV	Solenoid valve, 115 V AC coil
INT	Integrator, Bell & Howell 19-407A
S-H	Sample & Hold, Bell & Howell 20-419

Figure 48. Biological inhibitor detector---signal block diagram



Legend in Table 8

Table 8. LEGEND FOR SIGNAL BLOCK DIAGRAM OF FIGURE 48

Position of Selector Switch	Recorded Parameter
1	dissolved oxygen
2	endogenous equilibrium dissolved oxygen
3	integration of oxygen uptake curves of standard 1 and 2
4	area of oxygen uptake curve of standard 1 (A_1)
5	area of oxygen uptake curve of standard 2 (A_2)
6	$[(A_1)(\text{coefficient}) - A_2]$ comparator output signal recorded is A_1 (coefficient)

The operation of the automated cycle is as follows:

> Cam 1 activated a 0- to 15-second timer which opened a solenoid valve for a preset time interval to introduce 500 ml of the bacteria into the reaction vessel. Cam 6 switched on the mixer 15 seconds after filling the reaction vessel. The bacterial sample was aerated to an airflow rate of 36 ml/min and mixed at a rate of 200 to 400 rpm for a period of 16 minutes. This was enough time for the DO level to reach equilibrium at 80 to 90 percent of saturation, representing the constant endogenous oxygen uptake in equilibrium with the oxygen transfer from the air supply. The rate of mixing and airflow were kept constant throughout the measurement cycle. At this point, Cam 2 activated a three-way solenoid valve allowing 10 ml of the standard propylene glycol solution (600 mg/l) into the reaction vessel. It also activated a relay to start the integrator and caused sample and hold unit No. 1 to store the endogenous DO value. Cam 9 activated sample and hold unit No. 2 to sample the integrator 11 minutes after the standard sample was introduced. At the end of 12 minutes, Cam 3 opened a three-way solenoid valve on the feed sample line allowing 10 ml of the waste water feed into the reaction vessel. The feed was continuously flowing through the solenoid valve into a stand pipe which held 10 ml to the overflow level. After 18 minutes of exposure time (enough to degrade the feed samples), Cam 4 closed and repeated the operation of Cam 2 by adding another sample of the standard propylene glycol solution. Cam 10 activated sample and hold unit No. 3 to sample the integrator 11 minutes after the second standard sample was introduced. At the end of 12 minutes

Cam 5 opened the solenoid valve on the bacteria line to dump the contents of the reaction vessel, while Cam 6 turned off the mixer during the two-minute period that the reaction vessel was empty so that the mixer did not lose its magnetic coupling. The total cycle time was 60 minutes and the cam timer was a continuous mechanism by which the same cycle was repeated every 60 minutes.

The integrated area of the oxygen uptake curve of the first standard sample, which was stored in sample and hold unit No. 2, was multiplied by a preset fraction, usually 0.5, and compared to the integrated area due to the second standard sample, which was stored in sample and hold unit No. 3. The comparator operated a relay that set off an alarm or diverted the feed to a holding pond when the oxygen uptake of the second standard was less than 0.5 times the oxygen uptake of the first standard.

The instrument was designed for flexible operation and had many features for easy trouble shooting and maintenance. A recorder-selector switch located in front of the instrument allowed recording of any of six different outputs that are shown in Figures 48 and its legend (Table 8). A dump button on the front of the instrument actuated the solenoid valve on the bacteria sample line to drain the contents of the reaction vessel at any time for the purpose of routine cleaning of the reaction vessel or resetting the measurement cycle.

The electrical controls were quite versatile. The total measurement cycle time was varied by simply changing a gear in the 10-cam timer, and each cam switch was adjustable for duration and phasing.

The electronic components were also very versatile since the gain of each component was adjustable.

INSTRUMENT TESTING ON MINIPLANT OPERATION

The automated biological inhibitor detector was installed at the activated sludge miniplant to test its operation both under normal conditions and withtoxic substances addded to the feed to the plant.

The bacteria sample to the instrument was obtained from the aeration basin mixed liquor or the sludge recycle stream from the biosettler. The higher concentration of the biomass in the sludge recycle gave a higher initial slope of oxygen uptake, but the time required for recording the oxygen uptake curve was not significantly different from that obtained with the lower concentration of biomass in the aeration basin liquor. The sludge recycle sample was deficient in DO and required a much longer time of aeration to reach the equilibrium endogenous level, as compared with the aeration basin liquor. The biomass concentration in the aeration basin did not vary as much as that of the sludge recycle, and more reproducible oxygen uptake data were obtained with bacteria samples from the aeration basin.

Continuous operation of the instrument to monitor the oxygen uptake of fresh water bacteria under normal operating conditions for a period of three months showed a measurement precision of ±5 percent. The area of the oxygen uptake curve as integrated by the instrument was linear for the range of 200 to 1000 mg/l standard propylene glycol solutions. For the same concentration of viable biomass in the bacteria sample, the sensitivity of the oxygen uptake signal was dependent on the rate of mixing, airflow rate, and other factors that affect the overall oxygen transfer coefficient K_{La}. Sensitivity was varied by controlling the airflow rate and adjusting the gain of the electronic parts of the instrument.

The instrument was tested for the detection of toxic substances in the feed while it was continuously monitoring the activity ratio (A_2/A_1) of the activated sludge miniplant. Batch tests were conducted by adding varying concentrations of the toxic substance to the feed sample entering the reaction vessel of the instrument. Continuous feeding of a selected toxin in the feed was the basis of dynamic tests to evaluate the effects of accumulation of the toxin in the bacteria.

Batch Toxicity Tests

The same inorganic and organic toxic substances tested in the laboratory were used for testing the instrument while the miniplant was operated on fresh water-glycol feed and the high salt waste water from the glycol production plant.

The activity ratio, as recorded by the instrument, is plotted as a function of toxin concentration in Figures 49 through Figure 52. The effects of the inorganic toxins on the fresh water bacteria, Figure 49, were very similar to the resutls obtained in the laboratory tests, Figure 42, in which the order of toxicity was Hg > CN > ClO > Cu > Cd. In the case of the glycol-acclimated salt bacteria, Figure 50, the order of toxicity was CN > Cu > Hg > Cd. The less toxic effect of copper and its leveling off at concentrations higher than 10 mg/l in case of the fresh water bacteria may have been due to the multiplicity of the species as compared to the salt water acclimated bacteria, where one major species predominated.

Figure 49. Effects of inorganic toxins on activitv of alvcol-acclimated fresh water bacteria

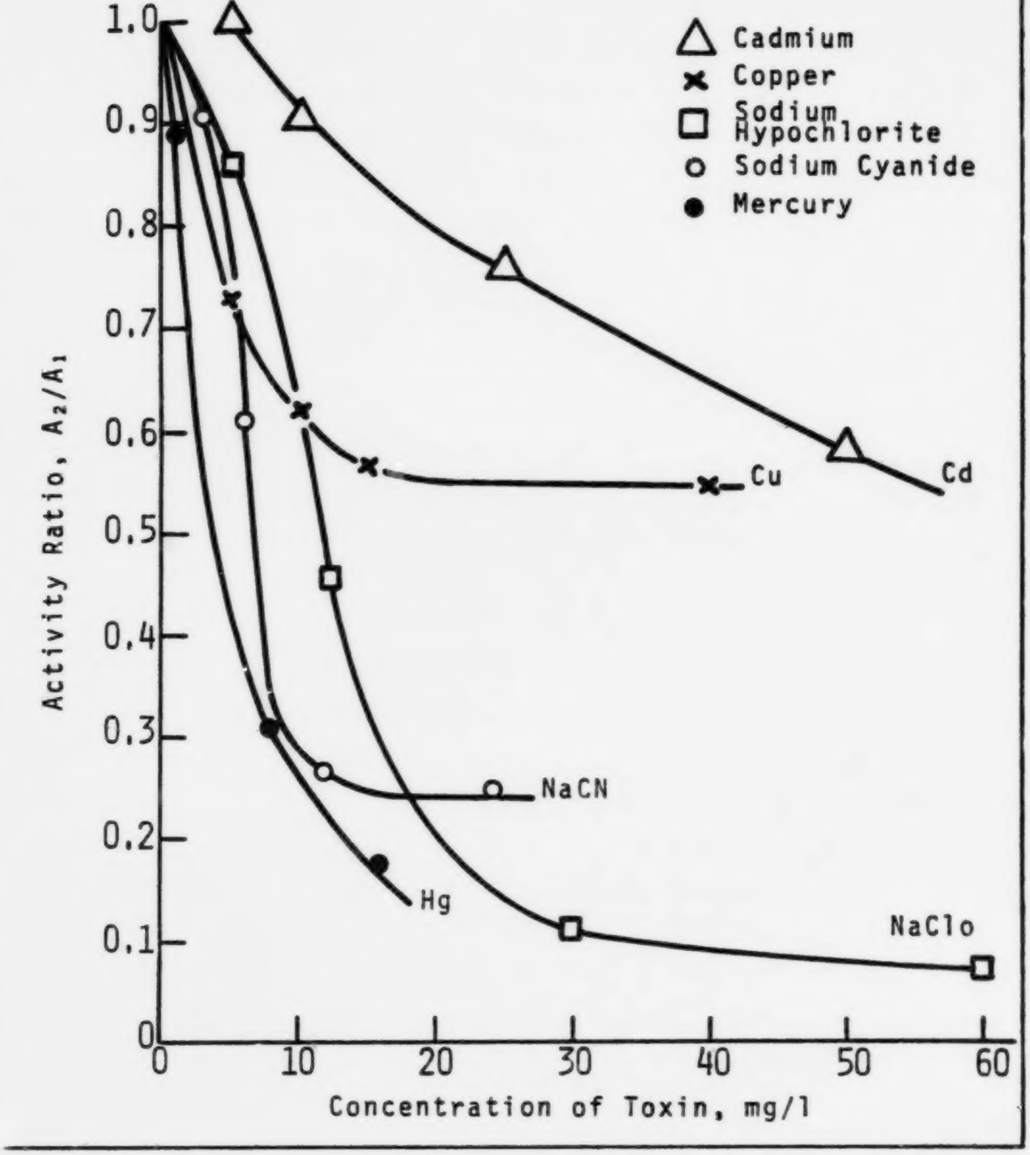
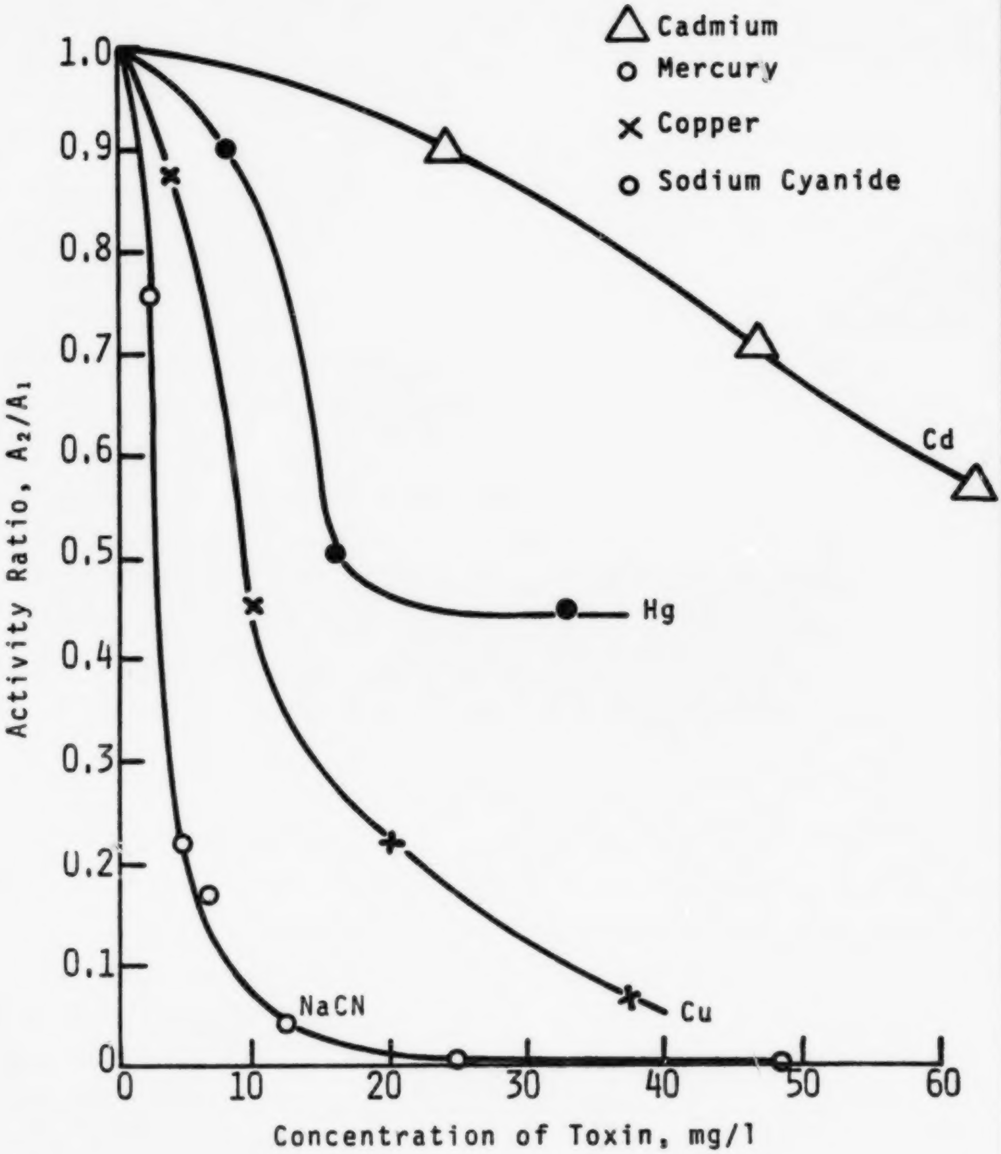


Figure 50. Effects of inorganic toxins on activity of glycol-acclimated salt water bacteria



Both phenol and dichloroisopropyl ether were quite toxic to the fresh water bacteria, Figure 51, while the ether had negligible effect on the salt water bacteria, Figure 52. This was clearly due to the presence of the ether in the waste water feed to the salt water bacteria, which was acclimated to concentrations of less than 100 mg/l of the ether.

Continuous Toxicity Tests

Dynamic toxicity tests were conducted by adding copper to the glycol waste water feed in the activated sludge mini-plant. The copper was added as cupric chloride to the equalization tank and fed continuously to the salt water bacteria. The area of oxygen uptake curve for a standard propylene glycol solution (A_1) was determined before the copper addition. The same area (A_2) was recorded by the ins-instrument once an hour after the addition of copper for a period of 24 hours. The activity ratio (A_2/A_1) for the test runs with 6.6 mg/l copper and 3.0 mg/l copper are shown in Figure 53 and Figure 54, respectively. The data scattering of Figure 53 was due to the malfunction of the stirring mechanism of the instrument; that was corrected during the second test.

Both tests showed no, or slight, reduction of activity for a period of six to nine hours followed by a sharp decline of the activity and reached a nil value after 16 hours and 21 hours for the 6.6 mg/l copper and 3.0 mg/l copper, respectively. A material balance of the copper during test B is shown in Table 9, and the concentration of copper in the bacteria and the liquid of the aeration basin are plotted as a function of time in Figure 55. During the 24 hours of feeding 3 mg/l of copper to the activated sludge

Figure 51. Effects of organic toxins on activity of glycol-acclimated fresh water bacteria

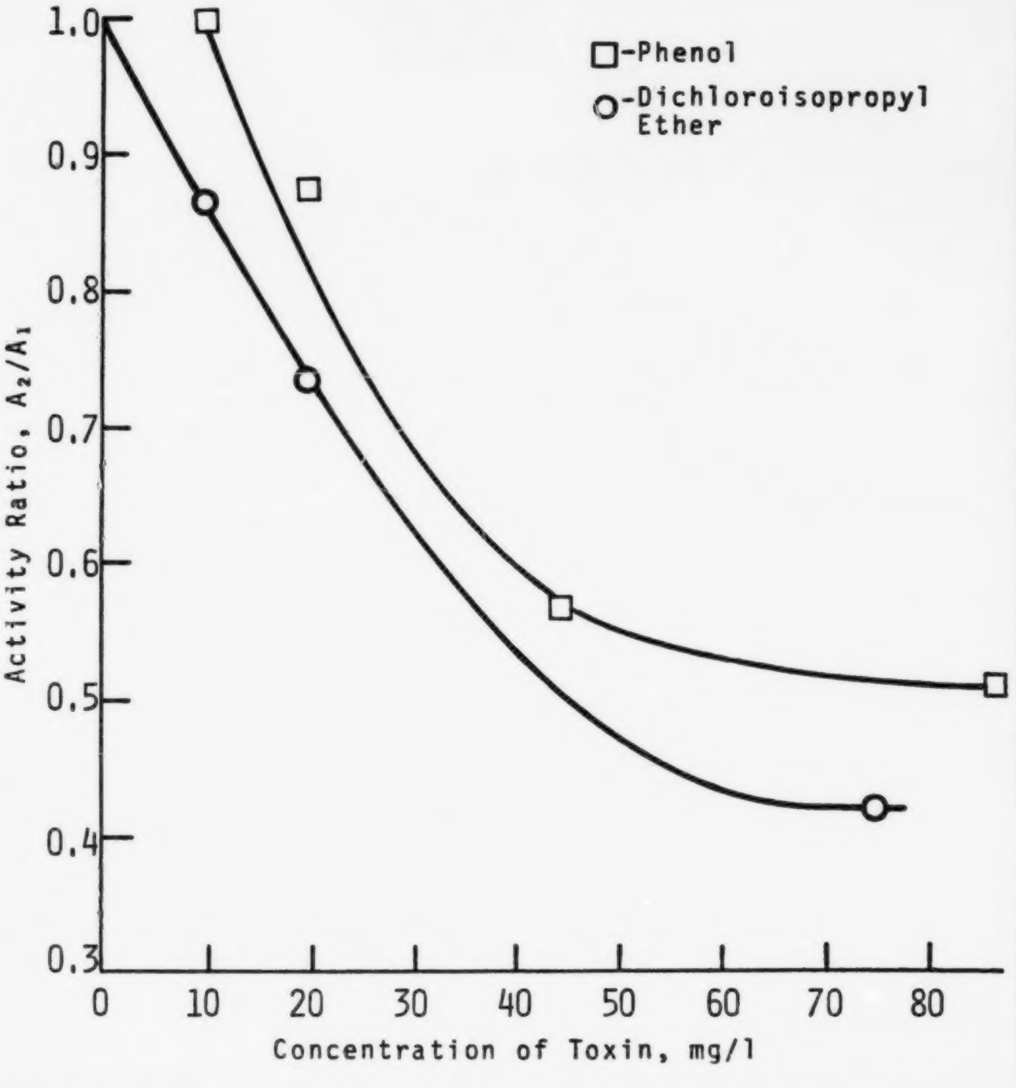


Figure 52. Effects of organic toxins on activity of glycol-acclimated salt water bacteria

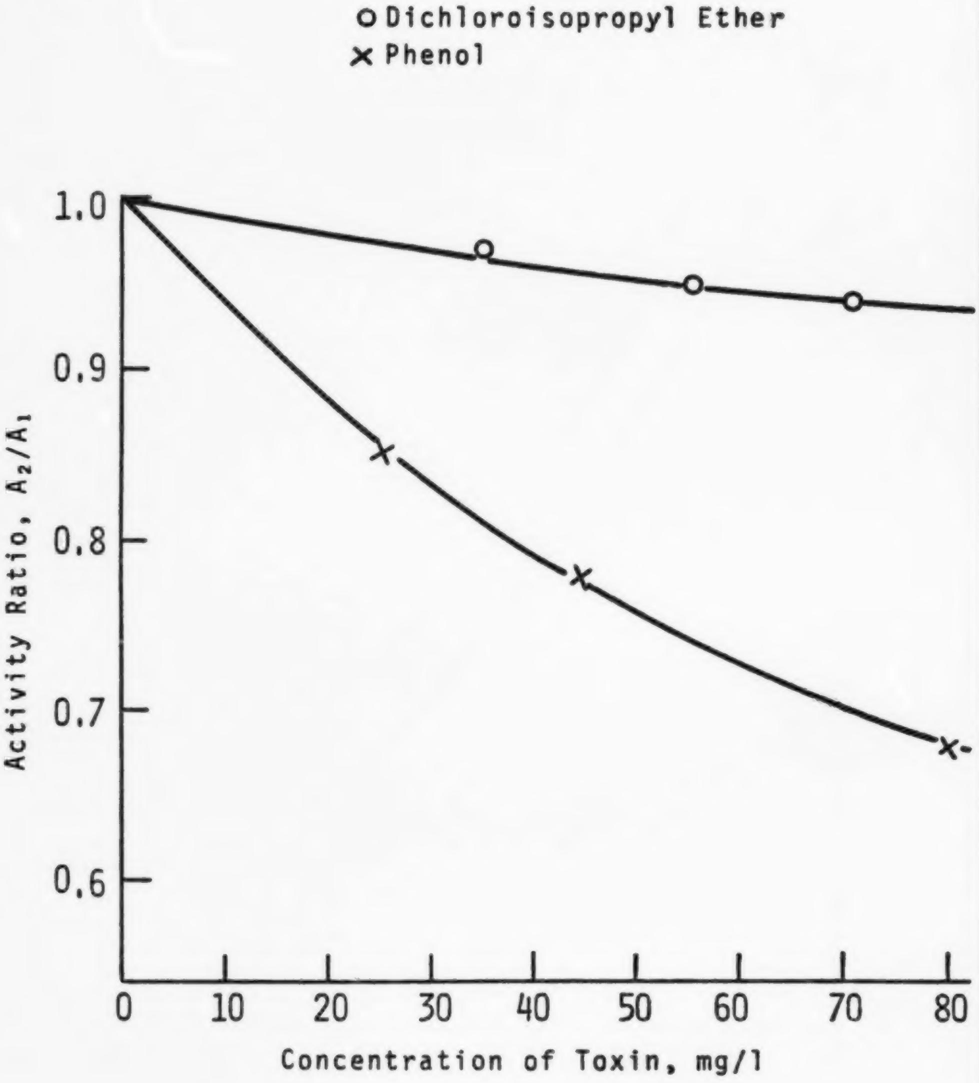


Figure 53. Dynamic toxicity test A, continuous feeding of copper to glycol-acclimated salt water bacteria

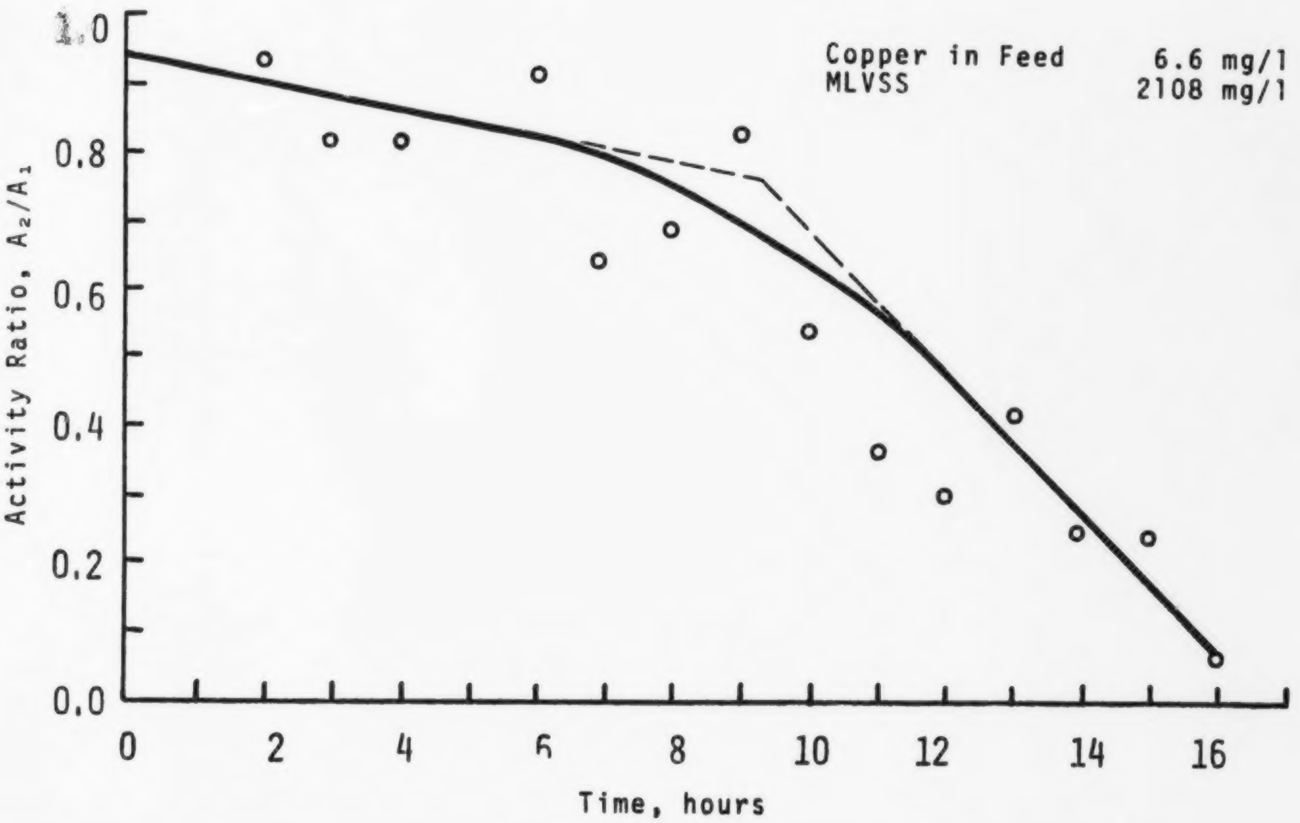


Figure 54. Dynamic toxicity test B, continuous feed of copper to glycol-acclimated salt water bacteria (copper concentration in feed = 3.0 mg/l bacteria concentration in feed = 1855 mg/l MLVSS)

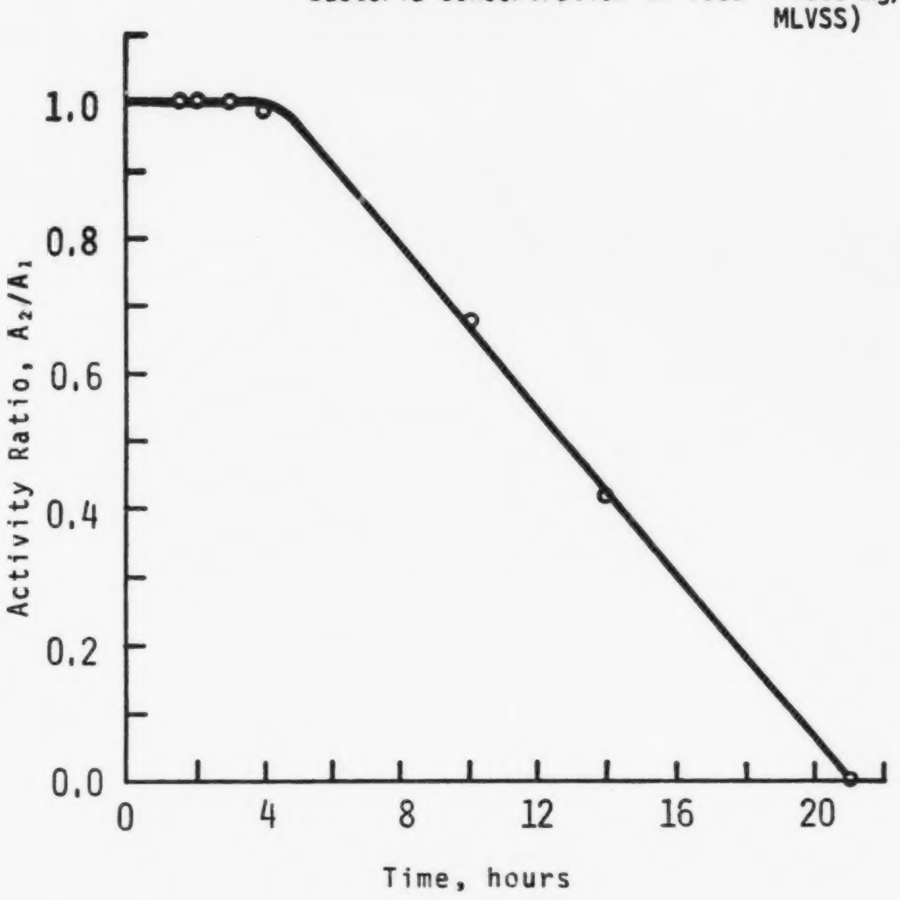


Table 9. DYNAMIC TOXICITY TEST B---
MATERIAL BALANCE ON COPPER
(BASIS 24 HOURS)

	Copper, mg
Copper In	10,044
Copper Out	
Overflow, liquid	2,253
Overflow, bacteria	956
Wasted, Bacteria	1,043
Total Out	4,252
In-Out	5,792

	Copper, mg
Copper Accumulation	
In Liquid	848
In Bacteria	5,048
Total Accumulation	5,896

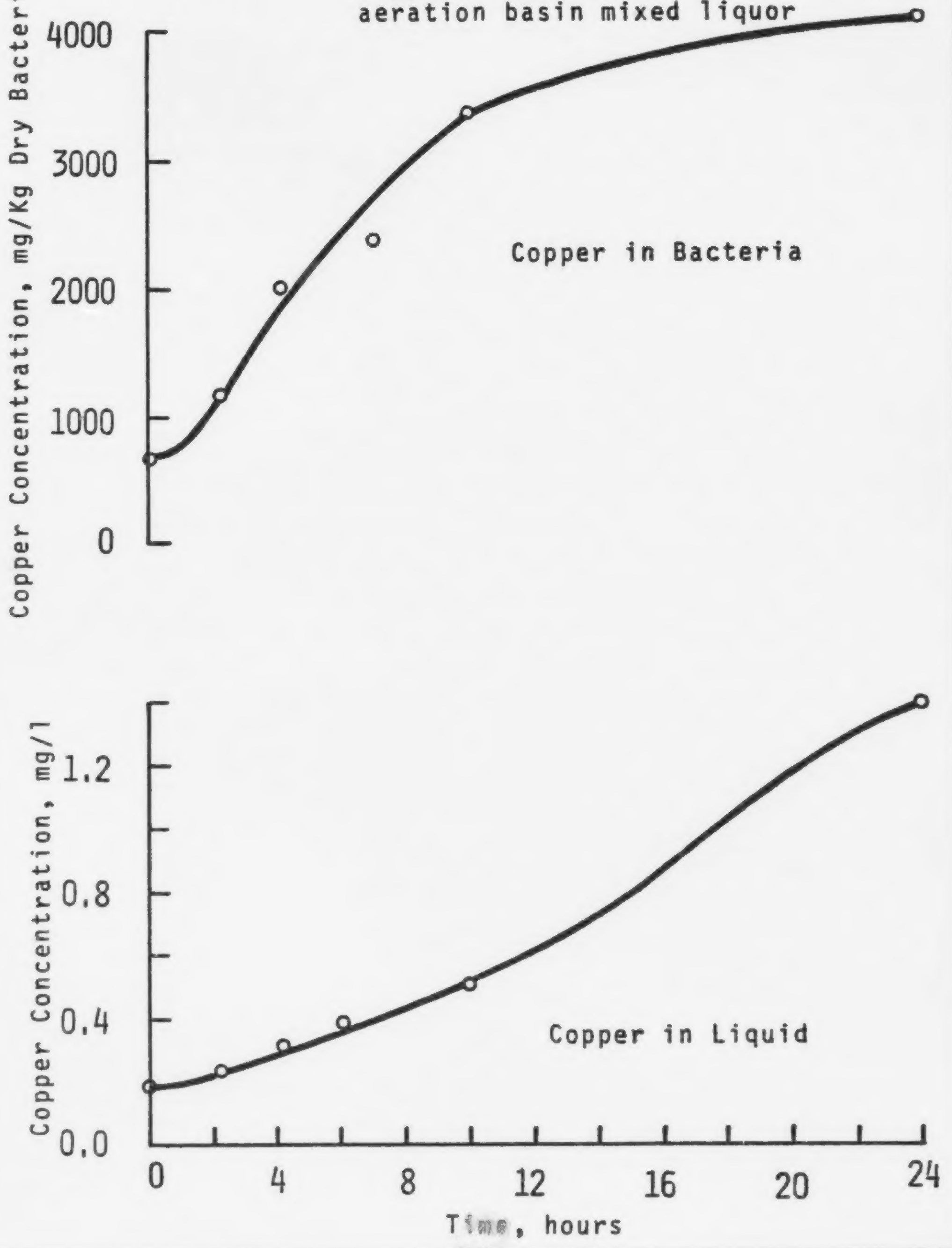
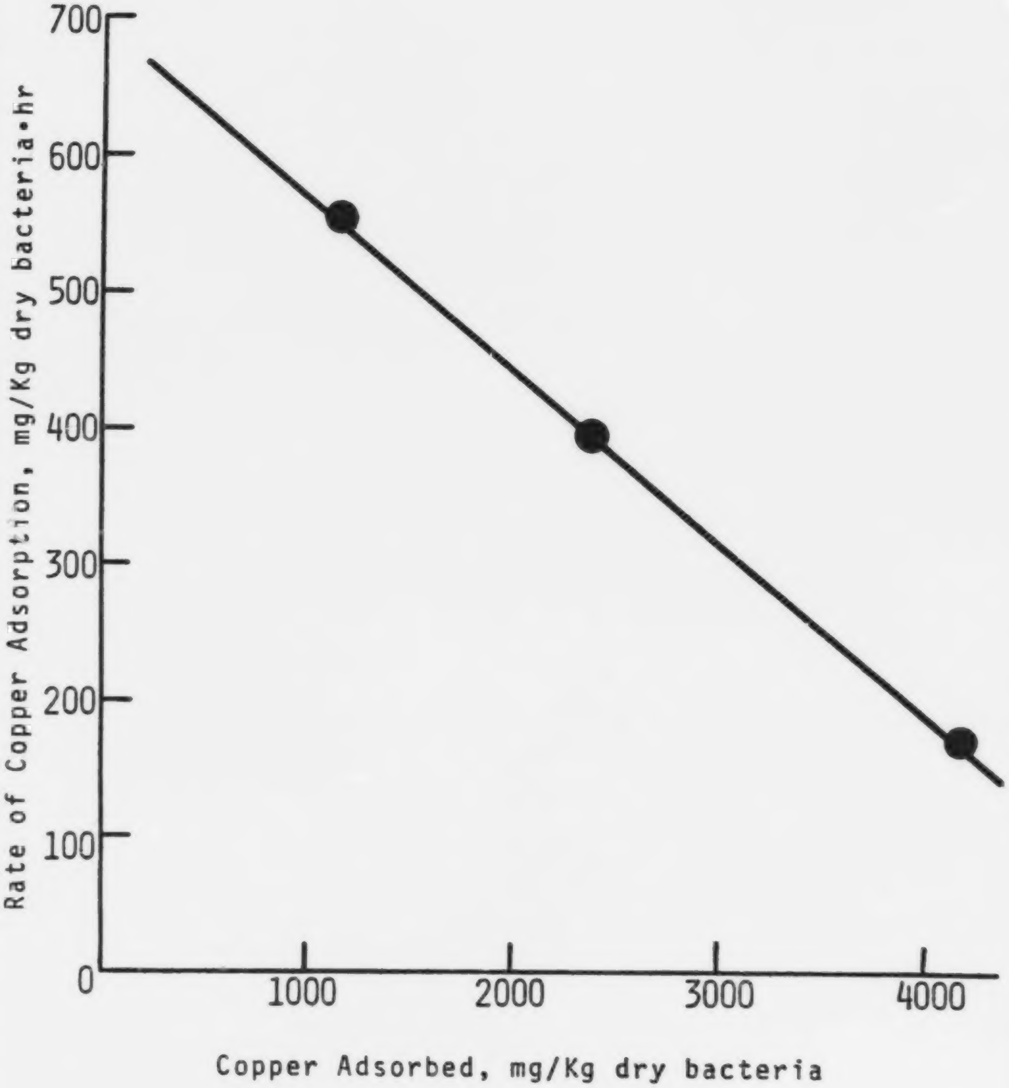


Figure 55. Dynamic toxicity test B, copper accumulation in aeration basin mixed liquor

miniplant, about 5800 mg of copper were accumulated in the bacteria, resulting in an increase of copper concentration from 656 to more than 4000 mg of copper per Kg of dry bacteria. The copper uptake rate by the bacteria was very rapid during the first eight hours, reaching a level of 3000 mg per Kg of dry bacteria. A plot of the rate of copper uptake versus the amount of copper adsorbed by the bacteria (Figure 56) indicates a second order reaction. This shows that the reaction with copper was controlled by the amount of respiratory enzyme available, which was large at the start of the test, showing no reduction in activity, then, at a point where most of the enzyme sites were complexed with copper, sharply decreased in activity.

Figure 56. Reaction order kinetics of copper adsorption on glycol-salt bacteria



SECTION IX

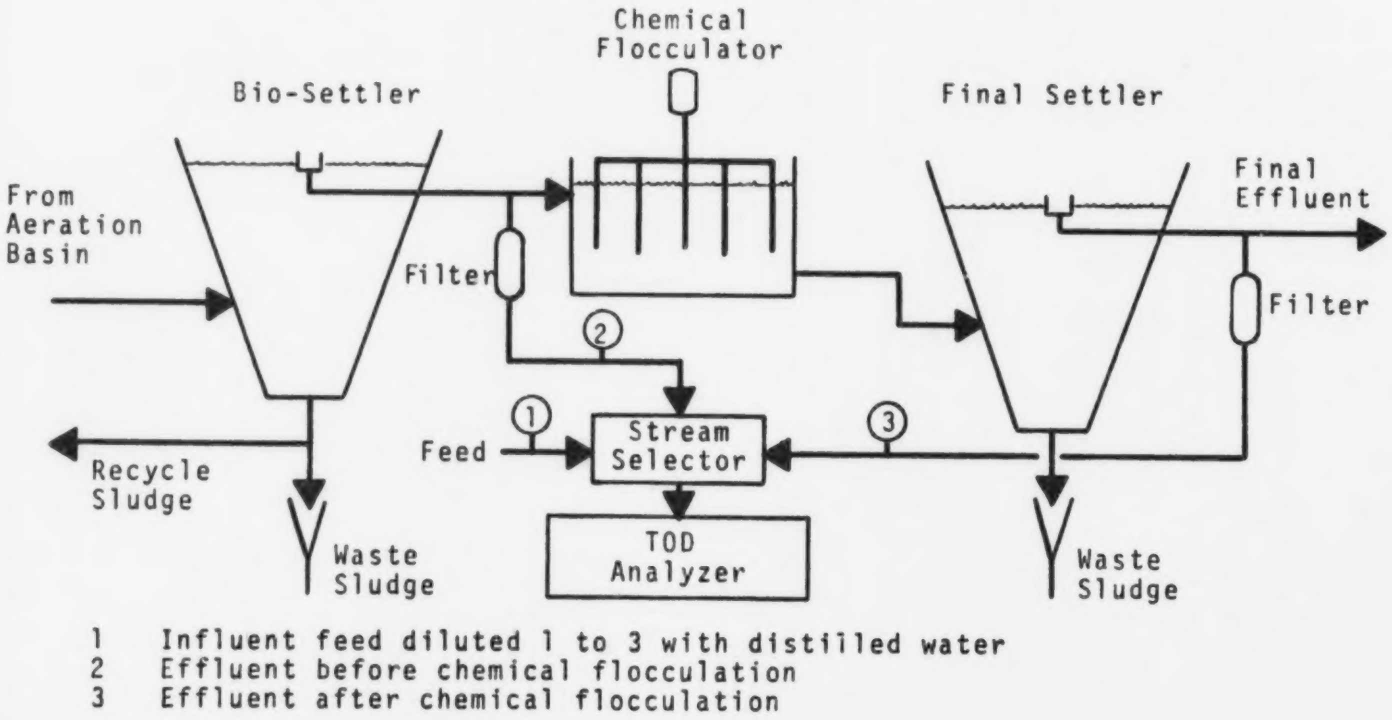
EFFLUENT QUALITY MONITORING AND INSTRUMENTS EVALUATION

An Ionics Model 225 Total Oxygen Demand (TOD) Analyzer and a multiple stream selector were used to monitor the feed to the activated sludge miniplant and the effluent before and after chemical flocculation.

A schematic of the monitoring system is shown in Figure 57. The five-stream selector was equipped with a four-stream manifold for the three streams to be monitored and for a standard solution used to calibrate the instrument response once a day. The feed stream was diluated 1:3 with distilled water using a cassette peristaltic pump. Gravity flow of the effluent samples to the instrument resulted in plugging the sample valves. Filters were added in the sample lines and pumping at a rate of 20 ml/min was necessary to maintain the sample flow to the instrument. The instrument was calibrated daily with a 700 mg/l TOD standard solution and the gain adjusted for a 70-percent full-scale response. The response that included the diluted feed and the effluent samples concentrations was linear in the range of 100 to 700 mg/l. The linear range of the calibration curve was checked once a week.

The Total Oxygen Demand Analyzer and the Total Carbon Analyzer, used for the control of nutrients addition and F/M control, were quantitatively evaluated for their performance during the periods of monitoring both fresh water and salt water streams (8 to 10 percent NaCl).

Figure 57. Schematic of effluent quality monitor



PERFORMANCE OF THE TOTAL OXYGEN DEMAND ANALYZER

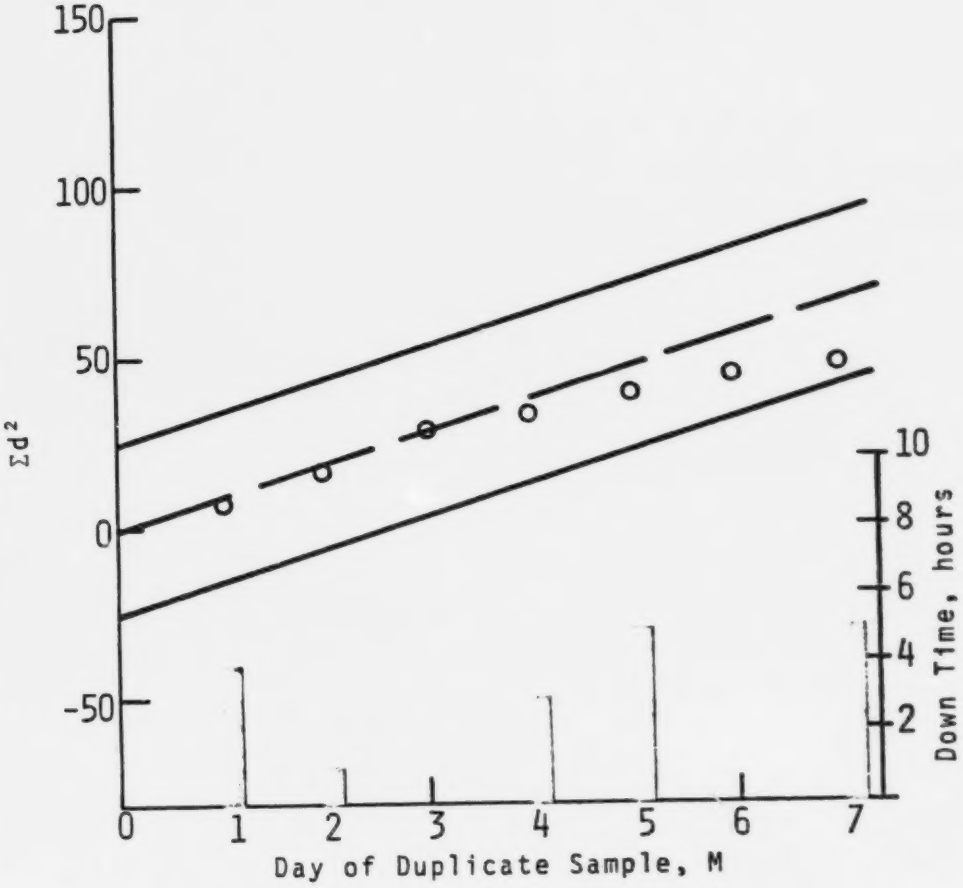
The instrument was operated in the laboratory for several months before being installed in the miniplant. The precision of measurement of standard solutions was ±4 percent in the linear range of 0 to 500 mg/l TOD. The rotary sampling valve required frequent adjustments to prevent leakage around the injection port and the drive gears for the motor turning the valve failed early during the second month of operation. The gears were replaced under warranty by the manufacturer at a cost of four hours labor and four days downtime.

The instrument and the multiple stream selector were installed in the miniplant and operated on fresh water samples for a period of four months and on salt water samples for one month. During the second week of operation, the printed circuit card in the stream selector, containing a stepping switch, failed and was replaced under warranty, resulting in three days downtime and four hours labor. This was followed by several problems with the instrument, a failure of a control card (two days of downtime), a failure of a power relay in the combustion heater circuit (two days downtime), and a failure of the balancing motor of the recorder (four days downtime). Numerous, short-duration downtimes occurred for cleaning the rotary sample valve and adjusting its O-ring injection seal. The oxygen sensor required adjustment of electrolyte level (that affects sensitivity) as frequently as twice a week and was disassembled for electrode cleaning once a month.

A precision chart and a record of the time out of control, downtime, and hours of maintenance labor were kept during the continuous operation of the instrument. A sample of the precision control chart is shown in Figure 58 and a summary

Figure 58. Precision control chart---
TOD analysis

Range - 45-65% Scale Full Scale = 1034 mg/l TOD
$\alpha = 0.05$ $\beta = 0.05$ $S_d = 1.90\%$ Scale



of the performance of the instrument is given in Table 10 for the fresh water and salt water sampling. The last month of operation on salt water (8 to 10 percent NaCl) samples proved to be very difficult. The combustion tube was coated with a salt glaze and carbon precipitated in the lines between the tube and the oxygen sensor, plugging the sample passage in less than four days of operation. The combustion tube was then slowly cooled to prevent its rupture and replaced every four days.

PERFORMANCE OF THE TOTAL CARBON ANALYZER

The instrument was operated in the laboratory for a period of three months, without incident. The precision of measurement of standard solutions was ±1 percent of the range of 0 to 300 mg/l organic carbon.

Operated on the miniplant feed stream was once interrupted by the failure of a heater element around the combustion tube that was replaced under warranty. For the purpose of monitoring the total carbon of a homogenized sample from the aeration basin, several controls were added to the analyzer. The added controls included a sampling switching circuit, a memory control, and a signal retransmission circuit. Operation of the analyzer on the homogenized bacterial samples caused frequent plugging of the injection valve until the sample handling system was improved to include a backwashing with the continuous feed sample. Replacement of the injection tube below the slide valve with one of larger bore resulted in significant loss of sensitivity.

A sample of the precision control chart is shown in Figure 59, and a summary of the instrument performance is given in Table 11 for the fresh water and salt water operation. During the four months operation on fresh water sampling one

Table 10. PERFORMANCE OF THE TOTAL OXYGEN DEMAND ANALYZER (Ionics Model 225-Serial No. 357)

	Fresh Water 4-mon. Per.	Salt Water 1-mon. Per.
Time in Control, %	62.5	52.6
Time Out of Control, %	35.4	41.6
Down Time, %	4.1	5.8
Maintenance Labor, hrs	95	20

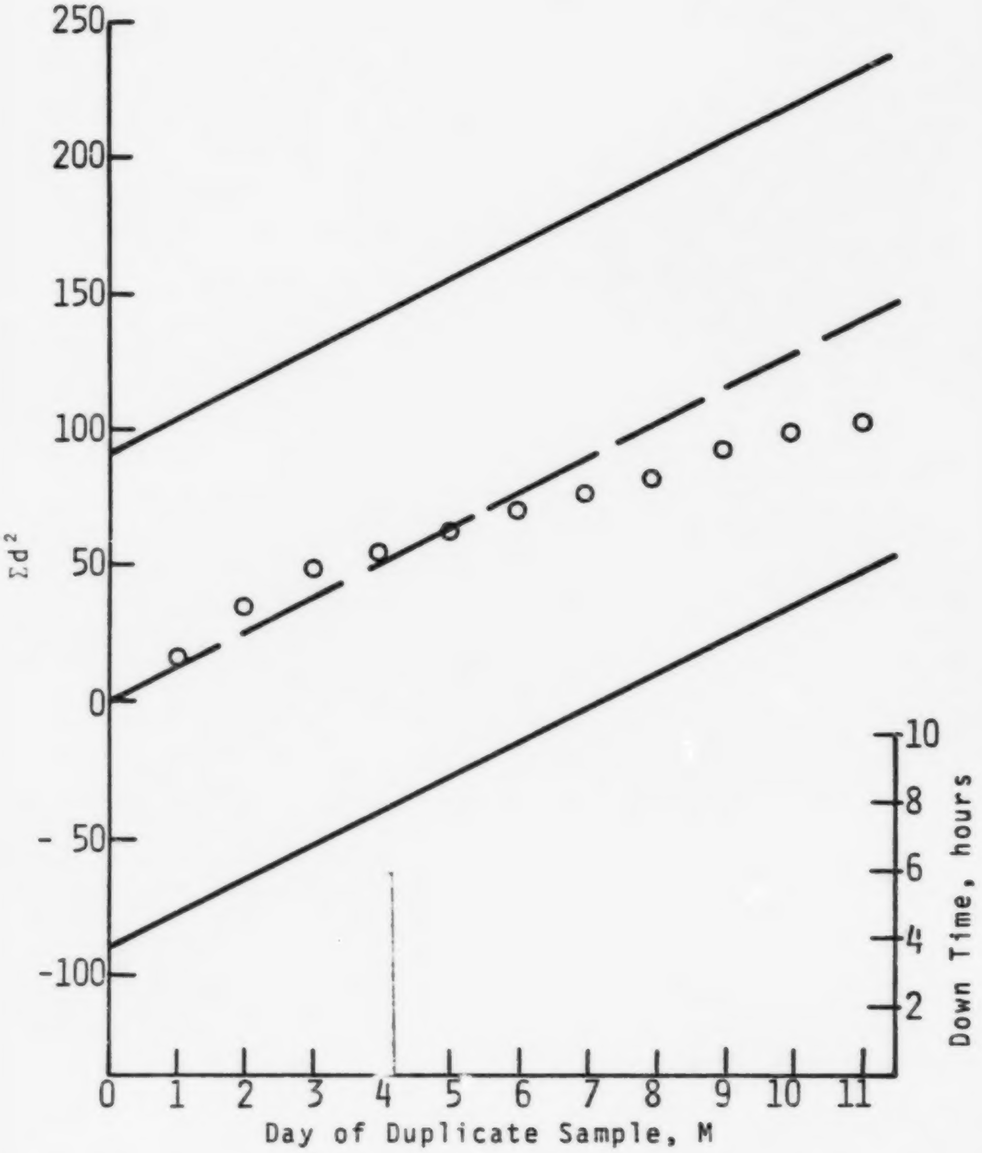


Figure 59. Precision control chart---
TC analysis

Range - 44-55% Scale Full Scale = 690 mg/l TC
$\alpha = 0.05$ $\beta = 0.05$ $S_d = \pm 3.67\%$ Scale

Table 11. PERFORMANCE OF THE TOTAL CARBON ANALYZER
(Ionics Model 1212-Serial No. 1196-12)

	Fresh Water 4-mon. Per.	Salt Water 1-mon. Per.
Time In Control, %	82.2	97.5
Time Out of Control, %	14.3	2.0
Down Time, %	3.5	0.5
Maintenance Labor, hrs.	60	2

incident resulted in most of the downtime and 50 hours of maintenance labor. The loss of sensitivity and the malfunctioning of the automatic zero electronic unit attached to the infrared detector resulted in a need to overhaul many parts of the instrument, including the infrared unit adjustment to restore its optical balance. Satisfactory operation was restored only after replacement of the combustion tube. Apparently, the catalyst inside the tube had fallen below the high temperature region and was the main cause of the erratic results.

Operation on the high salt (8 to 10 percent NaCl) sampled at 18-minute intervals was excellent for a period of three weeks followed by a slow loss of sensitivity to the end of the one-month operation. This was due to accumulation of salt within the combustion tube.

The slide injection valve was smoothed by grinding twice during the one-year operation of the instrument. Electrical maintenance was limited to replacement of one wire in the infrared unit, while the recorder and associated signal conditioning circuitry required no maintenance during the entire operating period.

Comparing the performance of the two analyzers, it was concluded that:

1. The slide injection valve of the TCA was more reproducible and required less maintenance to prevent leaking than the rotary valve of the TOD analyzer.

2. The infrared unit of the TCA was far more accurate and precise than the oxygen sensor of the TOD analyzer.

3. The Hastelloy combustion tube of the TCA was not subject to thermal fracture as was the case with the quartz combustion tube of the TOD analyzer.

4. Samples containing salt were difficult for both analyzers to handle. Somewhat better service was expected with the TCA.

SECTION X

PROCESS DESIGN CONSIDERATIONS

The development of on-line control systems for the activated sludge treatment process opens up many possibilities for flexible and innovative process design.

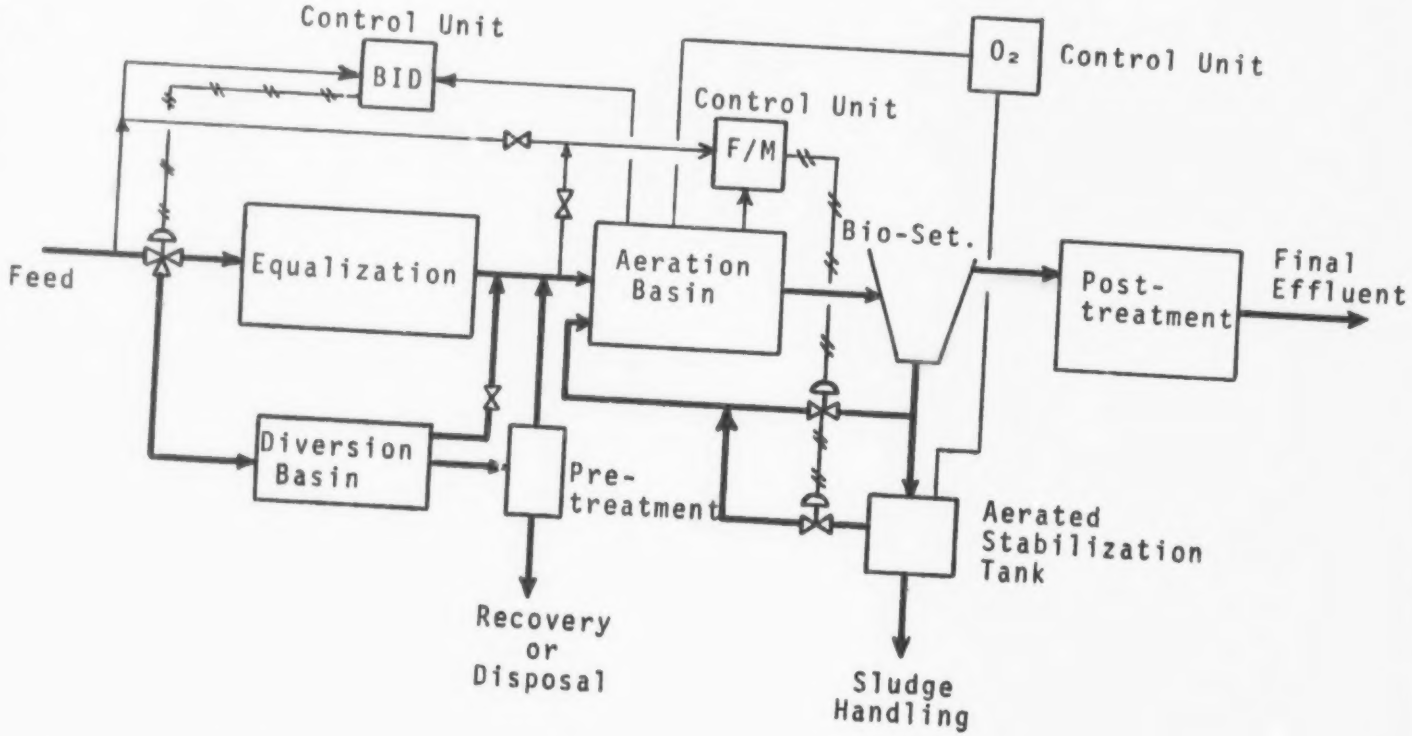
The two most important automatic control systems that could be incorporated in new plant designs are the biological inhibition detection (BID) system and the F/M control system. This is especially true for the design of facilities for the treatment of combined domestic and industrial wastes, where variations in the loading and the chemical composition of the waste stream may be very large. A typical domestic waste may have daily variations in BOD from 10 to 250 percent of the average value and flow rates may vary from 50 to 150 percent (Andrews, 1971). When combined with the possibility of slugs (or continuous flow) of industrial wastes where a toxin to the activated sludge biota may be present, the need for an inhibition detection system and an F/M control system becomes readily apparent.

Figure 60 illustrates a conceptual design of an activated sludge treatment system using these controls.

Equalization of the feed is very helpful for F/M control and is a necessary part of the plant design. Equalization dampens fluctuations in the organic content, in flow, in pH and in order physical-chemical characteristics of the feed to the aerated basin.

As illustrated in Figure 60, the equalization tank is also part of the feed-forward inhibition detection control system. Equalization provides a time delay, together with dilution of a possibly toxic feed during the biological activity analysis period.

Figure 60. Controlled activated sludge treatment



A toxic feed would ordinarily be diverted to a holding basin by the BID control system. After determination of the chemical nature of the toxin it could either be bled back into the system or treated by other means. Pre-treatment could include chelation of toxic metals and sorption of toxic organics.

The F/M control system would include measurement of flow and measurement of soluble organic concentration. The control system includes an aerated waste stabilization tank which supplies additional microorganism recycle in periods of high loading. The stabilization tank may be located as shown in Figure 60 or it may be located in the sludge recycle line. A pilot study would be useful in determining which site is more adequate in terms of microorganism viability and the effect on bio-flocculation in the aeration basin. For equal-sized stabilization tanks, a tank in the recycle line would give the shorter hydraulic residence time, which would indicate a higher biota viability. Having the tank located in the waste line would produce a more stabilized sludge, possibly aiding in bio-flocculation.

If the post-treatment step of chemical flocculation is used, another alternative in F/M control is possible. If the sludge from the floc-settler is viable, it may be used as the source of additional recycle in periods of high loadings. It has also been shown that both the treatment efficiency and the compactability of bulking sludge are improved significantly with the addition of aluminum hydroxide to the mixed liquor (Hsu and Pipes, 1973).

Two possible sources of the feed sample to be used for F/M control are shown in Figure 60. The location used for the experimental work in this study was in the feed line to the aeration basin (see Section V). Sampling from this location allowed easy proportional control.

If the feed sample for F/M control is taken before the equalization tank, F/M control will be more anticipatory in nature with very short F/M response times. The control system could then include a mini-computer which would be programmed to calculate the hydraulic effect that the equalization tank will have on the feed flow and composition and would affect F/M control on this anticipatory basis. Hence, extra sludge recycle could begin before the higher loadings affect the aerated biomass.

Another available control system that needs to be incorporated into the plant design is an aeration-basin oxygenation system. Besides optimizing oxygen allocations, the system should assure that proper agitation and micro-turbulence occurs so that the various mass transfer processes can be optimized (Kalinske, 1971). The supply of oxygen to the stabilization tank would also be controlled.

Nutrients and pH control would be provided for those processes requiring them. Tertiary treatment is necessary for colloidal and suspended solids removal from the effluent. Turbidity control would be provided for designs that employ flocculation as the tertiary treatment.

OPTIMUM DESIGN OF AN AUTOMATED ACTIVATED SLUDGE PLANT

The design of the aeration basin volume needed for a specified treatment efficiency is greatly affected by the incorporation of the F/M control system.

The choice of a steady-state model and the determination of the kinetic coefficients and other parameters pertinent to the chosen model are necessary in determining the aeration basin volume. A continuous pilot system is recommended for determining this data.

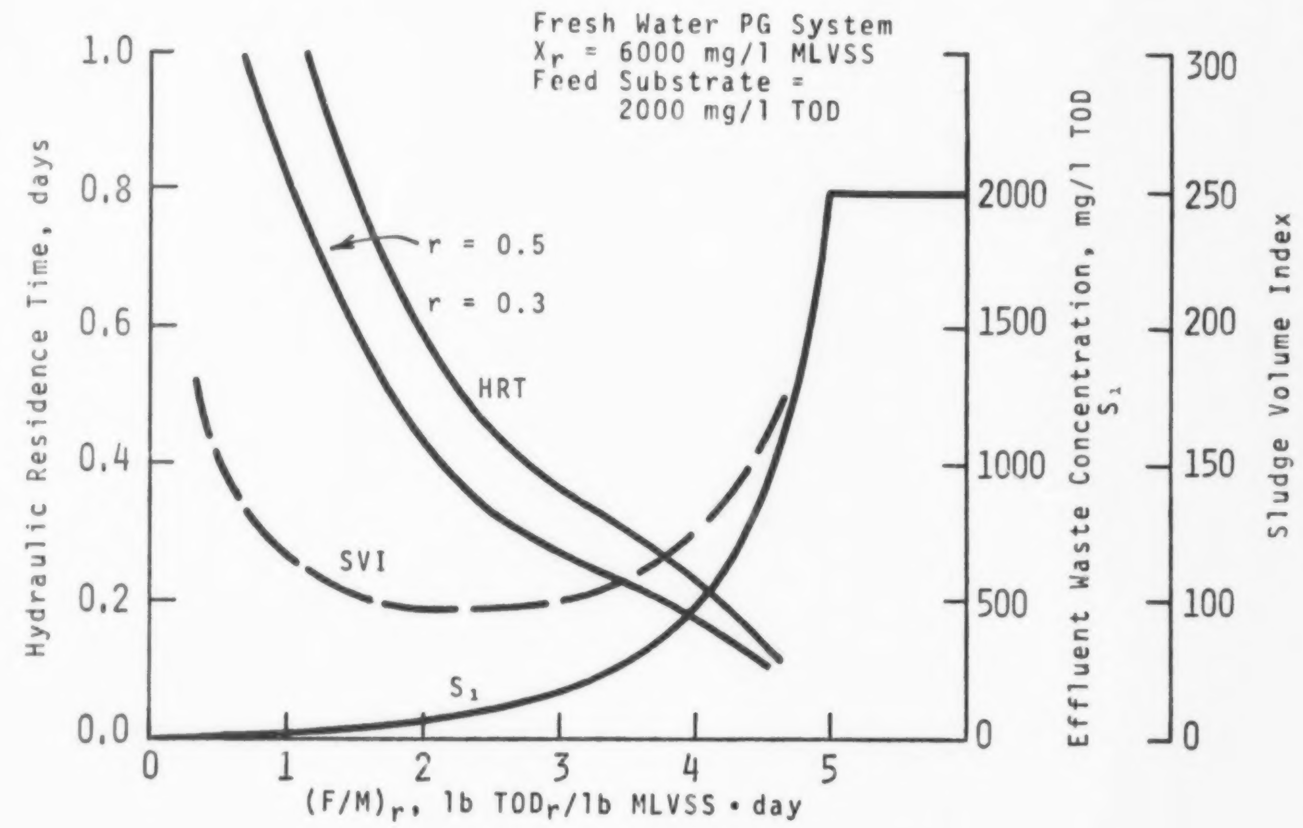
The steady-state model may then be used to calculate the required volume of the aeration basin as a function of other design variables. For example, a computer program using the Monod model steady-state calculations (completely mixed reactor) is presented in Appendix D and some of the design relationships for the fresh water, propylene glycol system are illustrated in Figure 61. The sludge compaction data (SVI) and the assumed recycle sudge concentration (X_r) are obtained from the pilot study. The treatment must be at an F/M within acceptable SVI limits. The aeration basin volume (the hydraulic residence time multiplied by the influent flow rate) is then a function of the recycle ratio.

The steady-state program also predicts the concentration of microorganisms in the aeration basin and the cell production rate, which at steady state is set equal to the wastage rate.

The aerated basin design should include evaluation of a transient-state model of the system to determine the effects that the suspected operational variations in loading will have on the process. The transient model permits optimum design of the F/M control system, both in terms of control action and stabilization tank volume. System response times can be determined to evaluate the adequacy of a given design.

The design should incorporate mixing and dispersion parameters obtained from actual equipment in order to insure that the residence time distribution in the constructed system will be

Figure 61. Steady-state design relationships



the same as was assumed for the designed system (Irvine, et al., 1973). This may entail further transient-state modeling to consider the possibility of a combination of two or more biological reactors.

The final design of the aeration basin volume is dependent on an economic evaluation of the alternative systems. Items to be considered include the reactor capital cost, recycle pumping costs, aeration and mixing costs, and control systems costs. The F/M ratio control costs include the stabilization tank capital and the cost of aeration. Benefits of stabilization, such as decreased sludge handling costs because of sludge digestion, must be taken into account. The benefits of increased process removals must also be weighed.

It is clear that the availability of developed control systems and of sufficiently accurate predictive modules will enable more efficient, more economic, and more controllable activated sludge systems to be built and to be operated at more closely regulated effluent limits. Sophisticated plants will require more sophisticated operation.

Future plant designs will have to incorporate the best available technology. Future plant operation will have to reflect the increased level of technology. Both design and operation depend upon the availability of sensors and the instrumentation necessary to use the sensed data to control the processes.

SECTION XI

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SECTION XII

LIST OF INVENTIONS

The following inventions are disclosed herein to the United States Government.

1. Biological Inhibitor Detector

The invention concerns the design and automation of a measurement cycle to detect the presence of toxic or inhibitory materials in waste water fed to a biological process. The measurement cycle is based on measuring the dissolved oxygen in a controlled biological reactor as a standard sample, a waste water feed sample, and another standard sample are consecutively added to the reactor. A toxicity index or activity ratio is calculated by comparing the oxygen uptake of the two standard samples before and after the bacteria sample has been exposed to the test sample. The invented measurement cycle is automated for repetitive monitoring of streams fed to a biological exidation process and can be used as a feed-forward control to divert a toxic feed before the loss or inhibition of the acclimated biomass.

2. Sampling System for a Homogenizable Solid-Liquid Mixture

Solid-liquid mixtures, such as an acclimated bacteria in an activated sludge aeration tank containing 2000 to 4000 mg/l suspended solids, are diluted as necessary and homogenized with a high-speed mixer. The homogenized sample then enters a stand pipe to be de-aerated, then flows by gravity to the sample injection valve of a monitoring instrument, and into the second leg of the U-shaped stand pipe. This flow flushes the sample valve, and upon injection of the homogenized sample all of the solenoid valves on the U-shaped stand pipe are de-energized to drain the excess sample. This restores the flow of a cotinuous sample of a clear waste water, e.g. the feed through the system, thus cleaning the lines from any residual from the batch sample.

SECTION XIII

GLOSSARY

ABBREVIATIONS

COD	chemical oxygen demand
DO	dissolved oxygen
JCU	Jackson candle turbidity units
MLTC	mixed liquor total carbon
MLVSS	mixed liquor volatile suspended solids
N:C	nitrogen to carbon ratio
SVI	sludge volume index
TC	total carbon
TC_F	total carbon of the feed
TC_M	total carbon of the mixed liquor
TCA	total carbon analyzer
TOD	total oxygen demand
TOD_r	total oxygen demand removed
VSS	volatile suspended solids

MATHEMATICAL SYMBOLS

A	waste calibration constant in equation 10
B	waste calibration constant in equation 10
C	concentration of DO in a bacteria sample, mass/volume
C_s	saturation DO concentration in a bacteria sample, mass/volume
E	percent efficiency of substrate removal, $100\ (S_0-S_1)/S_0$
F	food concentration, mass/volume
F/M	food to microorganisms ratio or loading
$(F/M)_r$	F/M "removed" or $(S_0-S_1)/X_{avg}$
k	substrate utilization kinetic constant--maximum rate of substrate utilization per unit weight of microorganisms, time^{-1}

k'	combined substrate biodegradation rate constant, volume/mass•time
k_d	microorganism decay coefficient, time^{-1}
K_e	endogenous reaction rate constant, time^{-1}
K_{La}	oxygen mass transfer coefficient, time^{-1}
K_s	half velocity coefficient, equal to the substrate concentration when the specific substrate utilization equals (1/2)k, mass/volume
M	microorganism concentration, mass/volume
η	bio-settler efficiency, percent
Q	influent waste flow rate, volume/time
R	recycle sludge flow rate, volume/time
r_r	oxygen uptake rate by microorganisms, mass/volume•time
S_0	feed soluble organic substrate concentration, mass/volume
S_1	soluble organic substrate concentration in aeration basin, mass/volume
T	total bio-settler underflow rate, volume/time
t	time
V	aeration basin volume
W	waste sludge flow rate, volume/time
W_{max}	maximum waste flow rate, volume/time
X	concentration of microorganisms in aeration basin, mass/volume
X_{avg}	average microorganism concentration in aeration basin over a chosen time interval, mass/volume
X_e	effluent microorganism concentration, mass/volume
X_r	recycle microorganism concentration, mass/volume
X	growth yield kinetic coefficient, mass microorganisms produced per mass of substrate utilized
Y	growth yield kinetic coefficient, mass microorganisms produced per mass of substrate utilized

SECTION XIV

APPENDICES

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APPENDIX A

OPERATION AND MAINTENANCE MANUALS

The automated activated sludge miniplant is operated unattended but the instrumentation and control systems require certain routine maintenance and trouble shooting practices that are detailed in this appendix. The operational procedures include some sampling collection and laboratory analyses to check the control systems and assess the overall performance of the process.

The details of the control systems applied to the operation of the aeration basin are shown in Figure 12. The automatic shutdown system is detailed in Figure 62. Operation and maintenance of these systems are described under the appropriate subsection of this appendix.

START-UP PROCEDURE

The influent flow to the equalization tank was started a day or two before plant start-up depending on the volume and residence time in the equalization tank. The stepwise start-up procedure was as follows:

1. Prepare the required concentration of chemicals and fill the alum, caustic, acid, and nutrients tanks. Turn the power on to heat the furnaces of the total carbon and total oxygen demand analyzers.

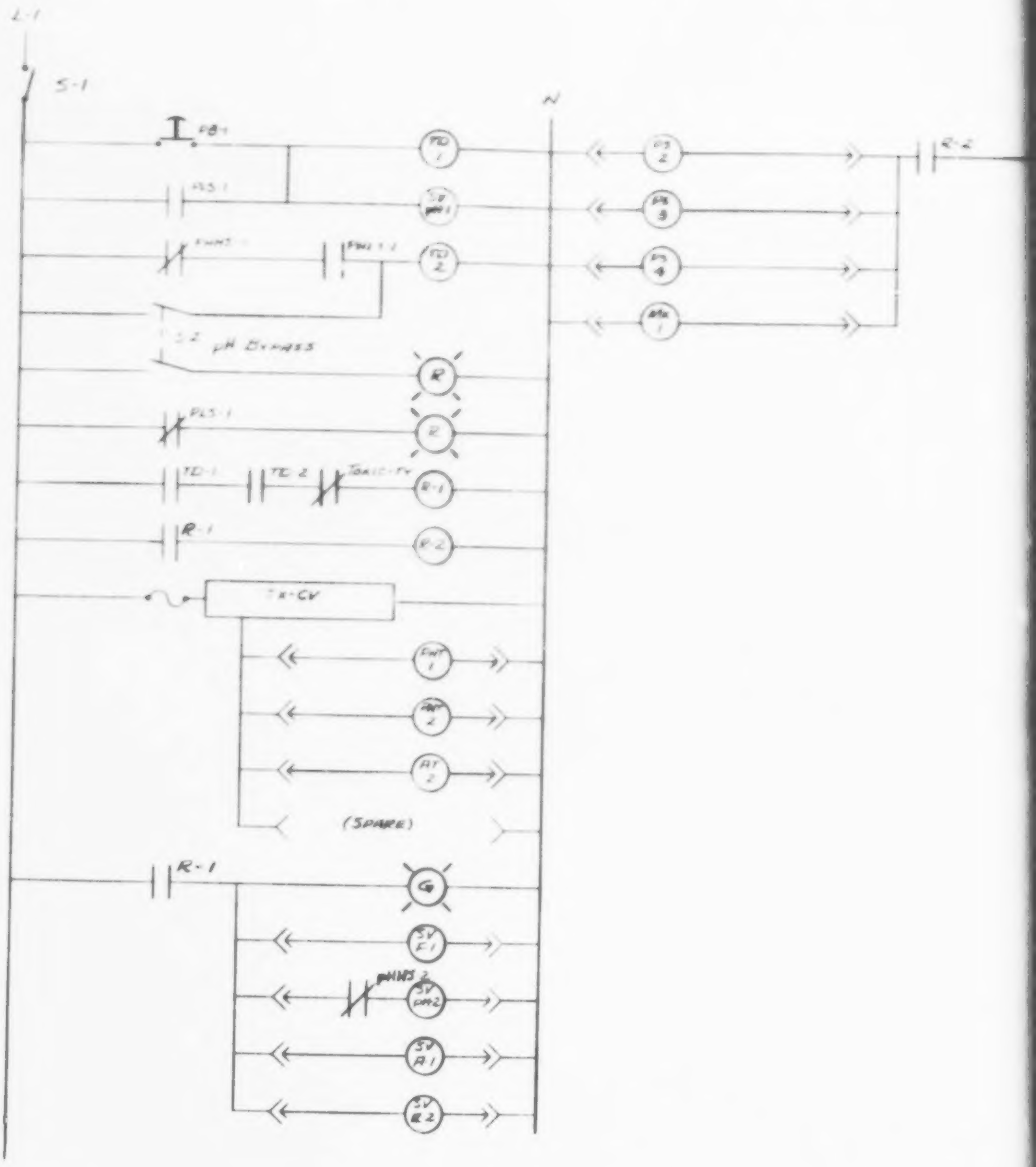
2. Add enough sludge to the aeration basin to obtain the desired mixed liquor suspended solids (MLSS) concentration when the basin is full. The source of the sludge could be from the sludge recycle of a sewage treatment plant or a batch-acclimated culture previously prepared

2. Adjusting the set points for the alum and caustic flow rates
3. Checking the liquid level of the alum and caustic tanks

The surface scatter turbidimeter required infrequent replacement of the electrical fuse, the lamp, and the meter relay lamp.

The Total Oxygen Demand Analyzer, monitoring the TOD in the feed and effluents required the following routine maintenance:

1. Changing filters to sampling system once a week
2. Checking rubber tubing in the peristaltic pumps
3. Checking leaks or plugging in sampling lines
4. Checking the rotary sampling valve for leaks or plugging
5. Checking the stream selector for proper operation sequence
6. Adjusting the nitrogen flow and electrolyte level in the oxygen detector of the instrument
7. Adjusting the level in the instrument scrubber, the detector current, and the automatic zero of the recorder for proper range of the signal
8. Replacing the scrubbing and detector cell solutions once a week
9. Adjusting sample valve O-ring clearance as necessary
10. Replacing furnace tube with freshly prepared catalyst at intervals found necessary by the operation
11. Changing the nitrogen cylinder when empty
12. Replacing burned out light bulbs



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REVISION

REVISION

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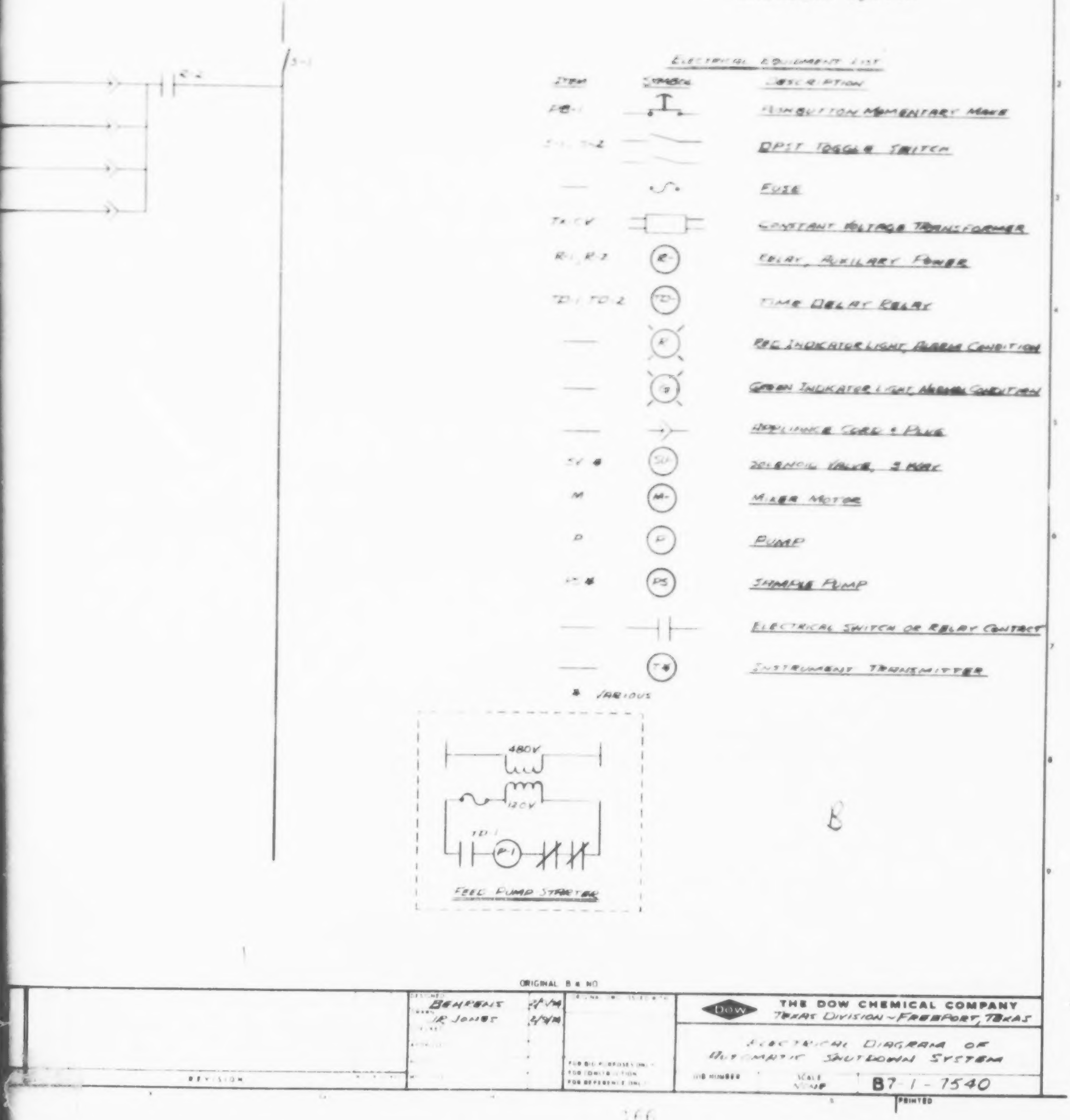
Figure 62

Electrical Diagram of Automatic Shutdown System



ELECTRICAL EQUIPMENT LIST

ITEM	SYMBOL	DESCRIPTION
PB-1		PUSHBUTTON MOMENTARY MAKE
S-1, S-2		DPST TOGGLE SWITCH
—		FUSE
TR-CV		CONSTANT VOLTAGE TRANSFORMER
R-1, R-2	R-	RELAY, AUXILARY POWER
TD-1, TD-2	TD-	TIME DELAY RELAY
—	R	RED INDICATOR LIGHT, ALARM CONDITION
—	G	GREEN INDICATOR LIGHT, NORMAL CONDITION
—		APPLIANCE CORD & PLUG
SV *	SV-	SOLENOID VALVE, 3 WAY
M	M-	MIXER MOTOR
P	P	PUMP
PS *	PS	SAMPLE PUMP
—		ELECTRICAL SWITCH OR RELAY CONTACT
—	T*	INSTRUMENT TRANSMITTER

* VARIOUS

in drums by a fill-and-draw procedure. Open the air supply valve and adjust the air rotameters to the required flow so as to maintain 2 to 3 mg/l dissolved oxygen in the aeration basin. A yellow springs oxygen meter is used to measure the DO in the aeration basin.

3. When the head in the equalization tank is enough to prime the feed-recycle pump, open all valves in the recycle loop. Turn the pH control on bypass and press the start switch. When recycle has started, set the pH at the control point midway between the upper and lower shutdown limits. Slowly close the ball valve on the pump suction until five inches of mercury vacuum is shown on the gauge. Open the acid valve and turn the pH controller to automatic to start the acid flow. Observe the rotameter to check the acid flow and when the pH, as recorded, is in control, turn the pH bypass off.

4. Start the feed to the aeration basin by turning the set point on the pneumatic proportional controller at the panel. This controller is calibrated for maximum flow of one gpm and is usually started at 0.5 gpm depending on the concentrations of the influent and MLSS in the aeration basin.

5. Check the temperature of the influent in the feed line as it enters the aeration basin. Control the temperature by turning on one or several of the electrical heaters around the feed line or diverting the feed through the cooling tower to maintain the temperature required for the test being conducted.

6. As the overflow from the aeration basin fills about half of the biosettler, start the sludge recycle pump. The variable drive on the pump is adjusted to the desired recycle flow rate. The rate is checked at the inlet to the aeration basin using a stopwatch and a one-liter graduated cylinder.

7. Observe the biosettler as it overflows to the flocculator and the last settling tank. Control the overflow weirs in the two settling tanks and remove the air trapped in the lines to obtain a smooth gravity flow through the flocculator and the last settling tank. The total overflow from the biosettler is directed through the Hach Surface Scattering Turbidimeter to the flocculator.

8. Put the Total Carbon Analyzer in service by turning on the recorder switch, the timer switch, and the stream selector switch. Keep the stream selector switch on Stream I (the feed) for a one cycle, then observe the instrument and the sampling system operation through a complete cycle. The sampling system includes Stream II (the bacteria sample) that is diluted, homogenized, and pumped through to the instrument slide valve.

9. Check the nutrient pump for proper operation and flow. The required flow rate of the nutrients to obtain the right ratio of organic carbon in the feed to the nitrogen and phosphorus in the nutrient solution is controlled by two variables. One is the speed of the nutrients pump that is adjusted manually. The other is the fraction of a (30 seconds) cycle that the pump is on. This is automatically controlled by the feed total carbon signal from the analyzer.

10. Operate the plant with total recycle and no sludge waste, bypassing the F/M control system, till the required steady-state concentration of MLVSS is reached. Adjust the controller, at the panel, that receives the F/M signal from the total carbon analyzer to obtain the required flow rate of sludge waste. This signal operates a three-way solenoid valve on the sludge recycle line and opens it to the waste line for a fraction of a 60-second cycle. The sludge waste flow rate is measured in a graduated cylinder at the wasted sludge drum. The maximum flow of wasted sludge is controlled by a limit switch that should be set so the sludge wasted during a 24-hour period would not exceed a precalculated maximum value. This limit switch guards against a washout of the bacteria in the event that the F/M control system fails.

11. Put the Hach Surface Scatter Turbidimeter in service by switching it on and setting the range to record the turbidity of the biosettler overflow. Open the manual valves on the alum and caustic tanks, and start the flocculator stirrer. The flow of the alum solution is controlled at the panel by adjusting the set point depending on the optimum dosage found by experiment. The flow of the caustic solution is controlled by setting the pH control system at the predetermined optimum pH (about 6.5).

12. Calibrate the Total Oxygen Demand Analyzer using standard solutions setting the instrument zero at 5.0 on the sensitivity dial. Put the stream selector in service by starting the sampling pumps and adjusting the stream

selector switches 1, 2, and 3. Check the flow of all samples and the signals on the recorder through one complete cycle. Readjust the sensitivity of the instrument to obtain a feed TOD signal around 75 percent of chart.

PLANT OPERATION, SAMPLING, AND ANALYSES

The operation of the automated miniplant was limited to daily and weekly routine checks of the control systems. The extent of sampling and analyses depended on the experimental design and the data required. The objective of the fresh water and salt water glycol feed tests was to evaluate the control systems. Accordingly, data collection was limited to those data necessary for this evaluation.

The influent pH-controller had a limit switch on high and low pH that closed the feed valve to the aeration basin if the pH got out of control and opened it when the pH control was regained. The control points were set at 5.5 and 9.0 with the set point at an intermediate value of 7.8. The pH electrodes were checked with buffer solutions as frequently as needed. A twice weekly check was found to be sufficient.

The nutrient control system was checked by running daily analysis of ammonia in the final effluent from the plant. The dial setting on the nutrient pump was then readjusted to minimize the residual ammonia in the effluent. A daily check of the amount of nutrients used was made by reading the calibrated sight glass on the nutrients tank, where each 3 mm was equal to one liter of solution. Using a nutrient solution of 1 percent ammonia and 0.1 percent phosphoric acid, normal operation of the plant consumed one tank of 55 gallons every two weeks of operation.

The food to microorganisms (F/M) control system required daily attention and service. The wasted sludge, collected in a 55 gallon drum, was measured and samples for analysis every 24 hours of operation. The sludge recycle rate was measured manually every day to check the performance of the recycle pump and to verify the recycle and waste sludge flow rates. The sampling system to the total carbon analyzer was cleaned daily to prevent any solids accumulation in the lines that could result in irregular analyses or even plugging of the sample flow.

The flocculation control system required very little attention. If the turbidity readings were recorded, the 24-hour results were obtained; otherwise, the indicator of the instrument should be read and entered in the data book daily.

The Total Oxygen Demand Analyzer, monitoring the feed and effluents, had no controls that affected plant operation. Calibration curves for this instrument and the Total Carbon Analyzer were run once a week. Daily, one standard was run to check of the two instruments, and the span control was reset if necessary.

Laboratory analyses were made daily as follows:

1. Feed grab sample: propylene glycol by VPC
2. Aeration basin mixed liquor: sludge volume index (SVI), Mixed liquor suspended solids (MLSS), and volatile suspended solids (MLVSS)
3. Biosettler overflow composite: MLSS, MLVSS, TOC, and ammonia

4. Final settler overflow composite: MLSS, MLVSS, and TOC

5. Sludge waste: MLSS and MLVSS

A plant data sheet was posted daily showing the data recorded by the instruments and the results of the laboratory analyses.

MAINTENANCE OF INSTRUMENTS AND CONTROLS

The instruments used for control and monitoring the operating parameters of the activated sludge miniplant required strict routine maintenance for good operation. Sample preparation and pretreatment was emphasized to minimize downtime for non-routine maintenance. Regular routine maintenance practices for recorders and support equipment are not covered in this text.

The influent pH-control system required the following routine and non-routine maintenance:

1. Inspecting and cleaning the pH-electrodes (once a week)
2. Calibration against high and low pH-buffer solutions (once a week)
3. Checking the acid line and rotameter for any leaks (daily)
4. Replacing defective electrodes (when necessary)
5. Cleaning acid flow rotameter (once a month)

The Total Carbon Analyzer (Ionics Model 1212) used to measure the TC in the feed and mixed liquor of the aeration basin required daily checking of the sampling system besides the following routine maintenance:

1. Checking the liquid level of the liquid gas separator
2. Checking the gas flow to the infrared analyzer
3. Adjusting wattmeter for correct operating wattage (185 watts)
4. Inspecting reaction chamber for oven heat
5. Checking the sample slide valve for leaks
6. Adjusting zero on the infrared analyzer
7. Back-flushing of sample flow to analyzer (once every two days)
8. Calibrating with one or two standards (daily)
9. Running all of the calibration curve (once a week)
10. Replacing nitrogen cylinder when empty
11. Replacing distilled water for water rinse (once a week)

In cases of out-of-control instrument response the main defective parts to check are:

1. Sample injection valve - which may have to be disassembled, cleaned, and reground to prevent leaking.

2. Reaction tube - which may have to be removed, and the catalyst and the injection tube cleaned.

3. Infrared detector - which may require removal and cleaning of the sample cell.

The flocculation control system, comprising the turbidimeter and the alum and caustic addition, required very little special maintenance:

1. Draining the turbidimeter once a day to clean the solids accumulated at the surface.

APPENDIX B

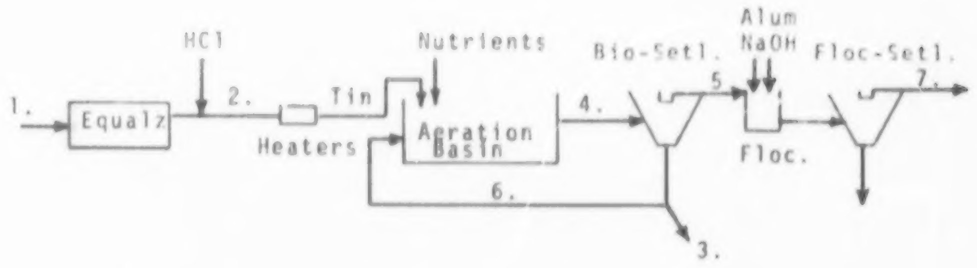
CALCULATIONS OF PLANT PERFORMANCE

A computer program was developed to facilitate routine data handling and to calculate parameters which describe the miniplant performance. The program was written in Fortran IV language and a Xerox Sigma 5 batch time-sharing monitor was used.

A sample data input form (#12) and the resulting computer printout displaying the summary of plant performance are presented. The logic of the program is also given (#13 and 14).

Definitions of symbols in the program can easily be related to the data input sheet and the corresponding READ statements and to the output printout and the corresponding WRITE statements. Material balances and parameters calculated have been described in the literature (Zeitoun, et. al., 1971).

Table 12. CALCULATIONS OF PLANT PERFORMANCE DATA INPUT FORM



M	D	Y
6	18	13

Flow to Equalz., GPM	Equalz. Vol, gal	D_1 lb/ft^3	TC_1 ppm	NaOH % wt	NaCl % wt	pH Eq.	T_2 °C
0.5	2580	62.4	473	0	0	6.5	

SV_{30} ml	Nutrient % NH_3	Nutrient % P	Nut. Flow ml/hr	HCl ml/min	$Turbid_5$ JCU
320	0.53	0.07	716		30

Flow to Basin, GPM	Tin °C	TC_2 ppm	D_2 lb/ft^3	pH_2 avg.	pH_2 high	pH_2 low	T_4 °C	Air Flow Readin
0.5	30	473	62.4	6.5	7.1	7.1	30	5.9

Basin Vol. gallon	$MLSS_4$ ppm	$MLVSS_4$ ppm	Yesterday's $MLVSS_4$ ppm	TC_4 ppm	D_4 lb/ft^3	pH_4
250	2188	2044	1776	836	62.4	6.5

TC_5 ppm	$MLSS_5$ ppm	$MLVSS_5$ ppm	D_5 lb/ft^3	Sludge Depths, inches Bio-set.	Sludge Depths, inches Floc-set.	T_5 °C	pH_5
41	15	13	62.4	14	30.5		

Alum ml/min	Caustic lb/hr	pH_7	$Turbid_7$ JCU	TC_7 ppm	D_7 lb/ft^3	$MLSS_7$ ppm	$MLVSS_7$ ppm	T_7 °C
11.8	0	6.5	2	19	62.4	0	0	

Recycle ml/min	Waste Sludge Depth, inches	$MLSS_6$ ppm	$MLVSS_6$ ppm	D_6 lb/ft^3	$Waste_3$ ml/min
885	16	5889	5818	62.4	92

TOD_2 ppm	TOD_5 ppm	TOD_7 ppm	PG_2 ppm	PG_4 ppm	PG_5 ppm	PG_7 ppm
1151	93	85	680			

TABLE 13. COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
1070 FORMAT(2X,'ECHO CHECK ON INPUT DATA'/)
 990 FORMAT(65X,'PAGE 3'//26X,'**SYSTEMS PERFORMANCE**'///
    C11X,'LOADING',14X,'REMOVAL'/,28X,'THRU CLARIFIER',9X,
    C'THRU FLOCCULATOR')
 989  FORMAT(7X,'LBS/DAY  LBS/DAY    LBS/DAY  LBS/DAY',
    C '  %WT  LBS/DAY  LBS/DAY  %WT')
 988 FORMAT(15X,'/LBMLVSS',12X,'/LBMLVSS',16X,'/LBMLVSS'/1X,
    C'TC',3X,F7.2,3X,F5.3,5X,F7.2,3X,F5.3,2X,F5.1,1X,F7.2,4X,
    CF5.3,2X,F5.1)
 987 FORMAT(30X,'CARBON BALANCE'/28X,'ON AERATION BASIN'/
    C16X, 'CARBON IN   CARBON OUT  CARBON TO CO2')
 986  FORMAT(17X,'LBS/DAY       LBS/DAY      LBS/DAY  %WT'/17X,
    CF7.2,5X,F7.2,4X,F7.2,2X,F4.1/)
 985 FORMAT(1X/,27X,'AERATOR PERFORMANCE'/34X,'AND'/
    C22X,'AERATION BASIN OXYGEN BALANCE'//11X,'O2 SUPPLIED',
    C8X,'O2 TRANSFERED          EFFICIENCY'/11X,'CFM   LB/HR',
    C7X,'KLA  CS*  LB/HR             %')
 984 FORMAT(9X,F5.2/)
 983 FORMAT(9X,'O2 CONSUMED',25X,'O2 CONSUMED'/
    C2X,'BASED ON CARBON BALANCE           BASED ON O2=',
    C'A(LBS TOD REM/DAY)+B(VSS)'2X,'LBS/DAY   LBS/LB TC',
    C' REM',15X,'LBS/DAY    LBS/LB TC REM')
 982 FORMAT(2X,F5.2,9X,F5.2)
 981 FORMAT(1X,'TURBTY  JCU',40X,F5.1,7X,F5.1//)
 949  FORMAT(6X,'FLOW  LBS/HR',4X,F8.1,19X,F8.1/12X,
    C'GPM',10X,F6.2,21X,F6.2)
 948  FORMAT(6X,'PH',18X,F4.1,23X,F4.1/6X,'TOC   PPM',
    C10X,F5.1,22X,F5.1/12X,'LBS/DAY'5X,F6.1,21X,F6.1)
 946 FORMAT(65X,'PAGE 2'//23X,'EQUILIZATION BASIN CONTENTS'//
    C20X,'TEMP',6X,'LIQUID INVENTORY HOLDUP TIME'/13X,'PH',
    C4X,'DEG C',10X,'GALLONS',10X,'HRS')
 945 FORMAT(12X,F4.1,4X,F5.1,9X,F7.1,10X,F5.1//)
 899 FORMAT(29X,'PH CONTROL SYSTEM'//4X,'EQUILIZATION BAS',
    C'IN',5X,'AERATION BASIN INFLUENT     HCL ADDITION'/
    C12X,'PH',13X,'HIGH PH   LOW PH  AVG PH',9X,'ML/MIN'/
    C10X,F4.1,14X,F4.1,6X,F4.1,4X,F4.1,9X,F6.1//)
 848 FORMAT(26X,'NUTRIENT ADDITION SYSTEM'//,1X,
    C'FLOW FROM          CARBON',9X,'NUTRIENT',14X,'RATIOS'/
    C1X,'EQUILIZATION   FROM EQUILZ.',3X,'FLOW   COMPOSITIO',
    C'N'/1X,'   LB/HR',9X,'LB/HR',6X,'LB/HR',3X,'%NH3',4X,
    C'%P',5X,'C:N',9X,'C:P'/3X,F7.2,5X,F8.3,6X,F5.3,2X,F5.2,
    C3X,F4.2,2X,'100:',F4.1,4X,'100:',F4.1//)
 845 FORMAT(28X,'F/M CONTROL SYSTEM'//2X,'WASTE FROM EQUIL',
    C'IZATION    AERATION BASIN    RECYCLE SLUDGE   WASTAGE',
    C/7X,'CONCENTRATION              MLVSS    TC       ',
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
  C'FLOW   MLVSS      FLOW'/3X,'GPM   PPM PG   PPM TC',
  C'       PPM      PPM        GPM    PPM',6X,'GPM'/2X,
  CF4.2,3X,F6.1,4X,F5.1,6X,F6.1,2X,F6.1,4X,F4.2,2X.F6.1,5X,
  CF4.2/)
844 FORMAT(22X,'FOOD/MICROORGANISM RATIO (LOADING)'/
  C1X,'FEED PG/',6X,'FEED TC/',6X,'LBS TC/DAY/',5X,
  C'LBS TC/DAY/LB MLVSS'/1X,'AER MLVSS',5X,'AER TC',8X,
  C'LBS MLVSS',4X,'BASED ON MLVSS=TC BUGS(113/60)'/1X,F4.2,
  C10X,F4.2,10X,F4.2,16X,F4.2//)
795 FORMAT(26X,'AERATION BASIN CONTENTS'//1X,'TEMP   PH',6X,
  C'AERATED LIQUID',6X,'MLSS',12X,'MLVSS',5X,'SLUDGE',
  C' AGE'/1X,'DEG C',10X,'LBS',6X,'GALLS    PPM    LBS',
  C'       PPM    LBS        DAYS')
794 FORMAT(1X,F5.1,1X,F4.1,3X,F8.1,2X,F7.1,2X,F6.1,
  C1X,F5.2,3X,F7.1,1X,F5.2,3X,F6.2)
793 FORMAT(1X,//1X,'RECYCLE      HOLDUP TIME BASED ON',
  C9X,'SLUDGE  TC REM RATE  TOD R.R.'/2X,'RATIO      IN',
  C'FLUENT    INFLUENT+RECYCLE   VOLUME   CONSTANT    C',
  C'ONSTANT')
792 FORMAT(4X,'%',9X,'HRS',14X,'HRS',9X,'INDEX',6X,'1/HRS',
  C5X,'1/HRS')
791 FORMAT(1X,F5.1,6X,F5.2,12X,F5.2,9X,F5.1,4X,F6.3,5X,F6.3/
  C//)
745 FORMAT(30X,'SETTLER DATA'//
  C20X,'OVERFLOW',10X,'INFLOW         BOTTOMS    SLUDGE'/22X,
  C'RATE',10X,'HOLDUP TIME     RATE',6X,'DEPTH'/15X,
  C'GAL/FT2/DAY   GPM',8X,'DAYS',9X,'GPM        INCHES')
695 FORMAT(1X,'BIO-SETTLER',5X,F6.1,5X,F4.2,5X,F6.2,9X,F4.2,
  C6X,F5.1/1X,'FLOC-SETTLER',4X,F6.1,5X,F4.2,5X,F6.2,9X,
  CF4.2,6X,F5.1///)
696 FORMAT(25X,'FLOCCULATION CONTROL SYSTEM'//17X,
  C'CAUSTIC      PH           ALUM',10X,'OVERFLOW'/18X,
  C'USUAGE',14X,'USUAGE',8X,'TURBIDITY'/19X,
  C'LB/HR',10X,'LB/HR',2X,'ML/MIN',8X,'JCU'/18X,F6.4,4X,
  CF4.1,1X,F6.4,F6.1,9X,F4.1/)
697 FORMAT(11X,'FLOCCULATOR INFLUENT     FLOC-SETTLER OVER',
  C'FLOW      % REMOVED'/18X,'LB/DAY',20X,'LB/DAY'/6X,'TC',
  C11X,F5.3,21X,F5.3,14X,F5.1/6X,'TOD',10X,F5.3,21X,F5.3,
  C14X,F5.1/)
645 FORMAT(1X,//,27X,'MICROORGANISM BALANCE'/29X,
  C'ON AERATION BASIN'/)
644 FORMAT(1X,'AVERAGE',14X,'VOLATILE SOLIDS FORMED',
  C12X,'RATIOS'/2X,'MLVSS',7X,'IN BASIN  WASTAGE',
  C'  CLAR OVERFLW   TOTAL    LBS VSS/DAY/'/1X,'PPM    LB',
  C'S',3X,'LBS/DAY',3X,'LBS/DAY',5X,'LBS/DAY',4X,'LBS/DAY',
  C3X,'MLVSS TOC REM')
643 FORMAT(1X,F6.1,F4.1,2X,F5.2,3X,F7.2,7X,F5.2,7X,F5.2,
  C5X,F4.2,2X,F4.2)
301 FORMAT(1X,'DENS LB/FT3',3X,F5.1,4X,F5.1,4X,F5.1,5X,F5.1,
  C3X,F5.1,4X,F6.1)
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
300 FØRMAT(65X,'PAGE 1'///15X,
   C'GLYCØL WASTE TREATMENT MINIPLANT A-1656'///
   C14X,'SUMMARY ØF PLANT',
   C' PERFØRMANCE FØR ',I2,1X,I2,1X,I2///25X,'STREAM',
   C' CHARACTERISTICS'/)
299 FØRMAT(13X,'WASTE TØ BASIN   BIØ-SETTLER  RECYCLE  WASTE
FLØC',
   C'-SETTLER'/14X,'BASIN   EFFLUENT ØVERFLØW     SLU',
   C'DGE  SLUDGE ØVERFLØW'/)
298 FØRMAT(1X,'FLØW  LB/HR ',F7.1,1X,F8.1,1X,F8.1,2X,F8.1,
   C1X,F7.1,2X,F8.1)
297 FØRMAT(9X,'GPM     ',F5.2,3X,F6.2,3X,F6.2,4X,F6.2,2X,
   CF6.2,4X,F6.2)
296 FØRMAT(1X,'TC       PPM    ',F5.1,13X,F5.1,23X,F5.1)
295 FØRMAT(6X,'LB/DAY ',F8.2,10X,F8.2,20X,F8.2)
294 FØRMAT(1X,'REMØVAL  %',22X,F5.1,23X,F5.1)
293 FØRMAT(1X,'TØD      PPM ',F7.1,12X,F6.1,22X,F6.1)
292 FØRMAT(6X,'LB/DAY ',F8.2,10X,F8.2,20X,F8.2)
290 FØRMAT(1X,'PH',13X,F4.1,42X,F4.1)
291 FØRMAT(1X,'REMØVAL  %',22X,F5.1,23X,F5.1)
289 FØRMAT(1X,'TEMP  DEG C',3X,F5.1,4X,F5.1,4X,F5.1,23X,
   CF5.1)
288 FØRMAT(1X,'PG',6X,'PPM',3X,F5.1,4X,F5.1,4X,F5.1,23X,
   CF5.1)
287 FØRMAT(1X,'MLSS     PPM',11X,F6.1,3X,F6.1,3X,F7.1,1X,
   CF7.1,4X,F6.1)
286 FØRMAT(6X,'LB/DAY',12X,F6.2,3X,F6.2,3X,F7.2,1X,F7.2,
   C3X,F6.1)
285 FØRMAT(1X,'MLVS     PPM',11X,F6.1,3X,F6.1,3X,F7.1,1X,
   CF7.1,4X,F6.1)
284 FØRMAT(6X,'LB/DAY',12X,F6.2,3X,F6.2,3X,F7.2,1X,F7.2,3X,
   CF6.1)
283 FØRMAT(1X,'RES      HRS',4X,F5.2,4X,F5.2,4X,F5.2,23X,
   CF5.2)
282 FØRMAT(1X,'NACL      %',5X,F4.1)
281 FØRMAT(1X,'NAØH      %',5X,F5.2)
278 FØRMAT(1X,'ALUM  LB/HR',51X,F5.3)
277 FØRMAT(1X,'BASE  LB/HR',51X,F5.3)
276 FØRMAT(1X,'SVI',20X,F5.1)
275 FØRMAT(1X,'SV MLS-30MINS',9X,F6.1///)
269 FØRMAT(1X,'TURB    JCU',21X,F5.1,23X,F5.1)
502 FØRMAT(3I)
500 FØRMAT(8F)
499 FØRMAT(6F)
498 FØRMAT(9F)
497 FØRMAT(7F)
496 FØRMAT(8F)
495 FØRMAT(9F)
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
  494 FORMAT(6F)
  448 FORMAT(7F)
  447 FORMAT(16X,'TO POND  FROM POND')
  446 FORMAT(1X,'TOD',2X,F7.2,3X,F5.3,5X,F7.2,3X,F5.3,2X,F5.1,
     C3X,F5.2,4X,F5.3,2X,F5.1///)
  442 FORMAT(1X,'SLUDGE AGE:',F6.2,'HRS.'/)
      READ(1,502)MTH,IDAY,IYR
      READ(1,500)HEWFLW,VOLCP,DENS1,TOC1,OH1,CHL1,PH1,TCP2
      READ(1,499)SV3,PERNH3,PERPHO,FLWNUT,HCLFLW,TURBT1
      READ(1,498)CPWFLW,TINBAS,TOC2,DENS2,PH2,PHABH,PHABL,TAB,
     CSMETER
      READ(1,497)VOLAB,SS4,VSS4,PDVSS4,TOC4,DENS4,PHAB
      READ(1,496)TOC5,SS5,VSS5,DENS5,SLDEPB,SLDEPF,TCL,PH5
      READ(1,495)FALUM,FNAOH,PHFL,TURBT2,TOC7,DENS7,SS7,VSS7,
     CTFL
      READ(1,494)CSWFLW,CWSLUG,SS6,VSS6,DENS6,WVFLW
      READ(1,448)COD2,COD4,COD6,PG2,PGAB,PG4,PG6
      SCWSLU=CWSLUG
C     CALCULATE AVERAGE RECYCLE AND WASTE FLOWS
      BOTFLW=WVFLW+CSWFLW
      CWSLUG=CWSLUG*1.64*3785/(24.*60.)
      CSWFLW=BOTFLW-CWSLUG
      HEVFLW=HEWFLW
      HEWFLW=HEVFLW*(60.0*DENS1*8.33)/62.4
      TOCF1=((HEWFLW*TOC1)/1.0E6)*24.0
      CPVFLW=CPWFLW
      CPWFLW=CPVFLW*(60.0*DENS2*8.33)/62.4
      TOCF2=((CPWFLW*TOC2)/1.0E6)*24.0
      RSTCP=VOLCP/(CPVFLW*60.0)
C     NUTRIENT ADDITION SYSTEM
      PDNUT=FLWNUT*8.33/(1000.*3.785)
      PDNH3=PERNH3*PDNUT*14./(17.*100.)
      PDPHO=PERPHO*PDNUT/100.
      TOCF2H=TOCF2/24.
      CP=TOCF2H/PDPHO
      CP=100./CP
      CN=TOCF2H/PDNH3
      CN=100./CN
C     AERATION BASIN
C     CHANGE MILLILITERS/MIN TO LB/HR
      SCSWFL=CSWFLW
      CSWFLW=CSWFLW*DENS6*60./(28.3*1000.)
      ABWFLW=(CPWFLW+CSWFLW)
      CWFLW=ABWFLW
      CSVFLW=(CSWFLW*62.5)/(8.34*DENS6*60.0)
      ABVFLW=CPVFLW+CSVFLW
      RR=(CSVFLW/CPVFLW)*100.
      RX=RR/100.0
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
191.000          AVVSS=(VSS4+PDVSS4)/2.
192.000          TOCI=(TOC2+(RX*TOC5))/(1.0+RX)
193.000          TOCFI=ABWFLW*TOCI*24.0/1.0E6
194.000          CODI=(COD2+(RX*COD4))/(1.0+RX)
195.000          CODFI=ABWFLW*CODI*24.0/1.0E6
196.000          WSS6=(CSWFLW*SS6*24.0)/1.0E6
197.000          WVSS6=(CSWFLW*VSS6*24.0)/1.0E6
198.000          WSSTB=WSS6
199.000          WVSSTB=WVSS6
200.000          SSTB=(WSSTB*1.0E6)/(ABWFLW*24.0)
201.000          VSSTB=(WVSSTB*1.0E6)/(ABWFLW*24.0)
202.000          CVFLW=(CWFLW*62.5)/(8.34*DENS4*60.0)
203.000          WSS4=(CWFLW*SS4*24.0)/1.0E6
204.000          WVSS4=(CWFLW*VSS4*24.0)/1.0E6
205.000          RSTAB=(VOLAB/(ABVFLW*60.0))
206.000          RSTABS=(VOLAB/(CPVFLW*60.0))
207.000          WTAB=VOLAB*8.34*DENS4 /62.5
208.000          SSAB=(WTAB*SS4)/1.0E6
209.000          VSSAB=(WTAB*VSS4)/1.0E6
210.000          AIRFLW=(SMETER-1.1)/4.36
211.000 C        F/M CONTROL
212.000          RPGVS4=PG2/AVVSS
213.000          FDARTC=TOC2/TOC4
214.000          TOC2LC=(TOCF2/(TOC4*VOLAB*8.33*DENS4/62.4))*(60./113.)*
215.000         C1E06
216.000 C        CLARIFIER
217.000          CWSLUG=(CWSLUG*DENS6/28.316)*60./1000.
218.000          FWFLW=CPWFLW-CWSLUG
219.000          FVFLW=(FWFLW*62.5)/(8.34*DENS5*60.0)
220.000          TOCF5=(FWFLW*TOC5*24.0)/1.0E6
221.000          WSS5=(FWFLW*SS5*24.0)/1.0E6
222.000          WVSS5=(FWFLW*VSS5*24.0)/1.0E6
223.000          CVSLUG=(CWSLUG*62.5)/(8.34*DENS6*60.0)
224.000          WSSS6=CWSLUG*SS6*24.0/1.0E6
225.000          WSVSS6=CWSLUG*VSS6*24.0/1.0E6
226.000          COVFLW=(FVFLW*60.*24.)/6.1
227.000          RSTC=50./(CVFLW*60.)
228.000          TOCRC=TOCF2-TOCF5
229.000          TOCPRC=(TOCRC/TOCF2)*100.0
230.000          BOTFLV=CSVFLW+CVSLUG
231.000 C        FLOCCULATOR
232.000 C        CHANGE ML/MIN TO LB/HR USING ALUM DENSITY 40,000MG/LITER
233.000          FALUMW=FALUM
234.000          FALUM=FALUM*60.*40./(454.*1000.)
235.000          TEWFLW=FWFLW
236.000          TEVFLW=TEWFLW*62.5/(8.34*DENS7*60.0)
237.000          TOCF7=(TEWFLW*TOC7*24.)/1.0E6
238.000          WSS7=(TEWFLW*SS7*24.0)/1.0E6
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
      WVSS7=(TEVFLW*VSS7*24.0)/1.0E6
      FVSLUG=(FWSLUG*62.5)/(8.34*DENS8*60.)
      FOVFLW=(TEVFLW*60.*24.)/6.1
      RSTF=50./(TEVFLW*60.)
      TOCRF=TOCF2-TOCF7
      TOCPRF=(TOCRF/TOCF2)*100.0
      FNAOH=FNAOH*60./454.
C
      SVI4=SV3*1000./SS4
      ABVSS=((VSS4-PDVSS4)*WTAB)/1.0E6
      CWVSS=WSVSS6+WVSS5
      DWVSS=(ABVSS+CWVSS)
      AVWVSS=(AVVSS*WTAB)/1.0E6
      SLDAGE=AVWVSS/(WVSS5+WSVSS6)
      TOC2L=TOCF2/AVWVSS
      RTOCC=(TOCRC/AVWVSS)
      RTOCF=TOCRF/AVWVSS
      RTOC=(TOC1-TOC5)/(RSTAB*TOC5)
      RCOD=(COD1-COD4)/(RSTAB*COD4)
      RSLUG1=DWVSS/AVWVSS
      RSLUG2=DWVSS/(TOCF2-TOCF5)
C CARBON AND OXYGEN BALANCE
      CIN=TOCF2
      COUT=(TOCF5+WSVSS6*0.53)
      DELTAC=CIN-COUT
      PCCO2=(DELTAC/CIN)*100.0
      RATEO2=(DELTAC*32.0/12.0)+(DWVSS*32.0/113.0)
      RO2TOC=RATEO2/TOCRC
      CODF2=COD2*CPWFLW*24.0*1.0/1.0E6
      CODF4=COD4*FWFLW*24.0/1.0E6
      CODF6=COD6*TEWFLW*24.0/1.0E6
      COD2L=CODF2/AVWVSS
      CODRC=CODF2-CODF4
      RCODC=CODRC/AVWVSS
      CODPRC=CODRC*100.0/CODF2
      CODRF=CODF2-CODF6
      RCODF=CODRF/AVWVSS
      CODPRF=CODRF*100.0/CODF2
      REMVTC=((TOCF5-TOCF7)/TOCF5)*100.
      REMVTD=((CODF4-CODF6)/CODF4)*100.
C     ECHO CHECK ON INPUT DATA
      WRITE(3,1070)
      WRITE(3,502)MTH,IDAY,IYR
      WRITE(3,500)HEVFLW,VOLCP,DENS1,TOC1,OH1,CHL1,PH1,TCP2
      WRITE(3,499)SV3,PERNH3,PERPHO,FLWNUT,HCLFLW,TURBT1
      WRITE(3,498)CPVFLW,TINBAS,TOC2,DENS2,PH2,PHABH,PHABL,
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
     CTAB,SMETER
      WRITE(3,497)VOLAB,SS4,VSS4,PDVSS4,TOC4,DENS4,PHAB
      WRITE(3,496)TOC5,SS5,VSS5,DENS5,SLDEPB,SLDEPF,TCL,PH5
      WRITE(3,495)FALUMW,FNAOH,PHFL,TURBT2,TOC7,DENS7,SS7,
     CVSS7,TFL
      WRITE(3,494)SCSWFL,SCWSLU,SS6,VSS6,DENS6,WVFLW
      WRITE(3,448)COD2,COD4,COD6,PG2,PGAB,PG4,PG6
      CALL PAGE(3)
      WRITE(3,300)MTH,IDAY,IYR
      WRITE(3,299)
      WRITE(3,298)CPWFLW,ABWFLW,FWFLW,CSWFLW,CWSLUG,TEWFLW
      WRITE(3,297)CPVFLW,ABVFLW,FVFLW,CSVFLW,CVSLUG,TEVFLW
      WRITE(3,296)TOC2,TOC5,TOC7
      WRITE(3,295)TOCF2,TOCF5,TOCF7
      WRITE(3,294)TOCPRC,TOCPRF
      WRITE(3,293)COD2,COD4,COD6
      WRITE(3,292)CODF2,CODF4,CODF6
      WRITE(3,291)CODPRC,CODPRF
      WRITE(3,301)DENS2,DENS4,DENS5,DENS6,DENS6,DENS7
      WRITE(3,290)PH2,PHFL
      WRITE(3,269)TURBT1,TURBT2
      WRITE(3,289)TINBAS,TAB,TCL,TFL
      WRITE(3,288)PG2,PGAB,PG4,PG6
      WRITE(3,287)SS4,SS5,SS6,SS6,SS7
      WRITE(3,286)WSS4,WSS5,WSS6,WSSS6,WSS7
      WRITE(3,285)VSS4,VSS5,VSS6,VSS6,VSS7
      WRITE(3,284)WVSS4,WVSS5,WVSS6,WSVSS6,WVSS7
      WRITE(3,283)RSTABS,RSTAB,RSTC,RSTF
      WRITE(3,282)CHL1
      WRITE(3,281)OH1
      WRITE(3,278)FALUM
      WRITE(3,277)FNAOH
      WRITE(3,276)SVI4
      WRITE(3,275)SV3
      WRITE(3,745)
      WRITE(3,695)COVFLW,FVFLW,RSTC,BOTFLV,SLDEPB,FOVFLW,
     CTEVFLW,RSTF,FVSLUG,SLDEPF
      CALL PAGE(3)
      WRITE(3,946)
      WRITE(3,945)PH1,TCP2,VOLCP,RSTCP
      WRITE(3,795)
      WRITE(3,794)TAB,PHAB,WTAB,VOLAB,SS4,SSAB,VSS4,VSSAB,
     CSLDAGE
      WRITE(3,793)
      WRITE(3,792)
      WRITE(3,791)RR,RSTABS,RSTAB,SVI4,RTOC,RCOD
```

Table 13 (continued). COMPUTER PROGRAM FOR CALCULATIONS OF PLANT PERFORMANCE

```
      WRITE(3,987)
      WRITE(3,986)CIN,COUT,DELTAC,PCCO2
      WRITE(3,985)
      WRITE(3,984)AIRFLW
      WRITE(3,983)
      WRITE(3,982)RATEO2,RO2TOC
      WRITE(3,645)
      WRITE(3,644)
      WRITE(3,643)AVVSS,AVWVSS,ABVSS,WSVSS6,WVSS5,DWVSS,
     CRSLUG1,RSLUG2
      CALL PAGE(3)
      WRITE(3,990)
      WRITE(3,989)
      WRITE(3,988)TOCF2,TOC2L,TOCRC,RTOCC,TOCPRC,TOCRF,RTOCF,
     CTOCPRF
      WRITE(3,446)CODF2,COD2L,CODRC,RCODC,CODPRC,CODRF,RCODF,
     CCODPRF
      WRITE(3,899)PH1,PHABH,PHABL,PH2,HCLFLW
      WRITE(3,848)CPWFLW,TOCF2H,PDNUT,PERNH3,PERPHO,CN,CP
      WRITE(3,845)CPVFLW,PG2,TOC2,VSS4,TOC4,CSVFLW,VSS6,CVSLUG
      WRITE(3,844)RPGVS4,FDARTC,TOC2L,TOC2LC
      WRITE(3,696)FNAOH,PHFL,FALUM,FALUMW,TURBT2
      WRITE(3,697)TOCF5,TOCF7,REMVTC,CODF4,CODF6,REMVTD
      CALL PAGE(3)
 20   STOP
      END
```

Table 14. SUMMARY OF PLANT PERFORMANCE---
June 18, 1973
Glycol Waste Treatment Miniplant
A-1656

STREAM CHARACTERISTICS

		WASTE TO BASIN	BASIN EFFLUENT	BIO-SETTLER OVERFLOW	RECYCLE SLUDGE	WASTE SLUDGE	FLOC-SETTLER OVERFLOW
FLOW	LB/HR	249.9	370.0	240.8	120.1	9.1	240.8
	GPM	.50	.74	.48	.24	.02	.48
TC	PPM	473.0		41.0			19.0
	LB/DAY	2.84		.24			.11
REMOVAL	%			91.6			96.1
TOD	PPM	1511.0		93.0			85.0
	LB/DAY	9.06		.54			.49
REMOVAL	%			94.1			94.6
DENS	LB/FT3	62.4	62.4	62.4	62.4	62.4	62.4
PH		6.5					6.5
TURB	JCU			30.0			2.0
TEMP	DEG C	30.0	30.0	.0			.0
PG	PPM	680.0	.0	.0			.0
MLSS	PPM		2188.0	15.0	5889.0	5889.0	.0
	LB/DAY		19.43	.09	16.98	1.29	.0
MLVS	PPM		2044.0	13.0	5818.0	5818.0	.0
	LB/DAY		18.15	.08	16.77	1.27	.0
RES	HRS	8.33	5.63	1.13			1.73
NACL	%	.0					
NAOH	%	.00					
ALUM	LB/HR						.062
BASE	LB/HR						.000
SVI			146.3				
SV MLS-30MINS			320.0				

SETTLER DATA

	OVERFLOW RATE GAL/FT2/DAY	GPM	INFLOW HOLDUP TIME DAYS	BOTTOMS RATE GPM	SLUDGE DEPTH INCHES
BIO-SETTLER	113.8	.48	1.13	.26	14.0
FLOC-SETTLER	113.8	.48	1.73	.00	30.5

Table 14 (continued). SUMMARY OF PLANT PERFORMANCE---June 18, 1973

EQUILIZATION BASIN CONTENTS

PH	TEMP DEG C	LIQUID INVENTORY GALLONS	HOLDUP TIME HRS
6.5	.0	2580.0	86.0

AERATION BASIN CONTENTS

TEMP DEG C	PH	AERATED LIQUID LBS	AERATED LIQUID GALLS	MLSS PPM	MLSS LBS	MLVSS PPM	MLVSS LBS	SLUDGE AGE DAYS
30.0	6.5	2081.7	250.0	2188.0	4.55	2044.0	4.25	2.95

RECYCLE RATIO %	HOLDUP TIME BASED ON INFLUENT HRS	HOLDUP TIME BASED ON INFLUENT+RECYCLE HRS	SLUDGE VOLUME INDEX	TC REM RATE CONSTANT 1/HRS	TOD R.R. CONSTANT 1/HRS
48.1	8.33	5.63	146.3	1.264	1.830

CARBON BALANCE ON AERATION BASIN

CARBON IN LBS/DAY	CARBON OUT LBS/DAY	CARBON TO CO2 LBS/DAY	CARBON TO CO2 %WT
2.84	.91	1.93	67.9

AERATOR PERFORMANCE AND AERATION BASIN OXYGEN BALANCE

O2 SUPPLIED CFM	O2 SUPPLIED LB/HR	O2 TRANSFERED KLA	O2 TRANSFERED CS*	O2 TRANSFERED LB/HR	EFFICIENCY %

O2 CONSUMED BASED ON CARBON BALANCE LBS/DAY	O2 CONSUMED BASED ON CARBON BALANCE LBS/LB TC REM	O2 CONSUMED BASED ON O2=A(LBS TOD REM/DAY)+B(VSS) LBS/DAY	O2 CONSUMED BASED ON O2=A(LBS TOD REM/DAY)+B(VSS) LBS/LB TC
5.67	2.18		

MICROORGANISM BALANCE ON AERATION BASIN

AVERAGE MLVSS PPM	AVERAGE MLVSS LBS	VOLATILE SOLIDS FORMED IN BASIN LBS/DAY	VOLATILE SOLIDS FORMED WASTAGE LBS/DAY	VOLATILE SOLIDS FORMED CLAR OVERFLW LBS/DAY	TOTAL LBS/DAY	RATIOS LBS VSS/DAY/ MLVSS	RATIOS LBS VSS/DAY/ TOC REM
1910.0	4.0	.56	1.27	.08	1.91	.48	.73

Table 14 (continued). SUMMARY OF PLANT PERFORMANCE---June 18, 1973

SYSTEMS PERFORMANCE

	LOADING		REMOVAL					
			THRU CLARIFIER			THRU FLOCCULATOR		
	LBS/DAY	LBS/DAY /LBMLVSS	LBS/DAY	LBS/DAY /LBMLVSS	%WT	LBS/DAY	LBS/DAY /LBMLVSS	%WT
TC	2.84	.714	2.60	.654	91.6	2.73	.686	96.1
TOD	9.06	2.279	8.52	2.144	94.1	8.57	2.156	94.6

PH CONTROL SYSTEM

EQUILIZATION BASIN	AERATION BASIN INFLUENT			HCL ADDITION
PH	HIGH PH	LOW PH	AVG PH	ML/MIN
6.5	7.1	6.1	6.5	.0

NUTRIENT ADDITION SYSTEM

FLOW FROM EQUILIZATION	CARBON FROM EQUILZ.	NUTRIENT			RATIOS	
		FLOW	COMPOSITION			
LB/HR	LB/HR	LB/HR	%NH3	%P	C:N	C:P
249.90	.118	1.576	.53	.07	100: 5.8	100: .9

F/M CONTROL SYSTEM

WASTE FROM EQUILIZATION			AERATION BASIN		RECYCLE SLUDGE		WASTAGE
	CONCENTRATION		MLVSS	TC	FLOW	MLVSS	FLOW
GPM	PPM PG	PPM TC	PPM	PPM	GPM	PPM	GPM
.50	680.0	473.0	2044.0	836.0	.24	5818.0	.02

FOOD/MICROORGANISM RATIO (LOADING)

FEED PG/ AER MLVSS	FEED TC/ AER TC	LBS TC/DAY/ LBS MLVSS	LBS TC/DAY/LB MLVSS BASED ON MLVSS=TC BUGS(113/60)
.36	.57	.71	.87

FLOCCULATION CONTROL SYSTEM

CAUSTIC USAGE LB/HR	PH	ALUM USAGE		OVERFLOW TURBIDITY JCU
		LB/HR	ML/MIN	
.0000	6.5	.0624	11.8	2.0

	FLOCCULATOR INFLUENT LB/DAY	FLOC-SETTLER OVERFLOW LB/DAY	% REMOVED
TC	.237	.110	53.7
TOD	.537	.491	8.6

The intput data needed for the execution of this program are:

CARD NO. 1

V - aeration basin volume (gallons)
Q - influent feed flow rate (gpm)
SFEED - initial influent feed concentration, (mg/l TOD)
SFEED2 - influent feed concentration after step change (mg/l TOD)
RECY - recycle flow rate (ml/min)
W - waste flow rate (ml/min)
XR - return sludge concentration (mg/l MLVSS)
KSUBS - half velocity coefficient (mg/l TOD)
YIELD - growth yield coefficient (lb VSS formed/lb TOD removed)
KENDOG - microorganism decay coefficient (day^{-1})

CARD NO. 2

XSTEP - time at which feed concentration is changed (hours)
XF - time at which simulation ends (hours)
H - stepsize (hours)
INT - frequency of data output desired (real-time data output every H*INT hours)

CARD NO. 3

Y(1) - initial substrate concentration in aeration basin (mg/l TOD)
Y(2) - initial microorganism concentration in aeration basin (mg/l MLVSS)

CARD NO. 4

FACTOD - conversion factor (mg TC/mg TOD)
FACVSS - conversion factor (mg TC/mg MLVSS)

CARD NO. 5

LOPT - simulation option: enter 0 for no F/M control
enter 1 for F/M control

AA - constant for waste calibration (see Equation 10)

BB - constant for waste calibration (see Equation 10)

WMAX - maximum waste flow rate allowable, (ml/min)

Mini-plant kinetic data was obtained in terms of MLVSS and TOD. Since the F/M control action was in terms of total carbon, it was necessary to have conversion factors FACTOD and FACVSS. The theoretical FACTOD conversion factor for propylene glycol is 0.28. However, the number entered into this program should be the actual conversion factor based on the operation of the TC and TOC instruments. Similarly, FACVSS should be the conversion factor from MLVSS to MLTC using values from on-line total carbon analysis (FACVSS is theoretically equal to 0.53 for cellular formula $C_5H_7O_2N$.)

The program and a sample printout are given in Tables 15 and 16, respectively.

APPENDIX C

TRANSIENT STATE ACTIVATED SLUDGE COMPUTER SIMULATION

A computer program was developed in order to predict transient changes in microorganism concentration, F/M and aeration basin substrate concentration to step changes in loading. The program logic uses the Monod type kinetic equations which are incorporated into the unsteady-state material balance equations (Equations 8 and 9). The F/M control is included in the program logic.

The computer program consists of the main program SIMUL, the logical function RKG, and the subroutine FXY.

The main program includes all the input (READ) and output (WRITE) statements. It also includes a calculation of the initial steady-state microorganisms concentration (XSTST) and the initial aeration basin substrate concentration (SONE) by using Monod kinetics. When inputting plant data, this calculation shows how close the process is to the theoretical steady state. The equations used are discussed in the literature (Lawrence and McCarty, 1970).

The material balance differential equations are solved in logical function RKG by using a fourth-order Runge-Kutta method with Gill's coefficients (Gill, 1951). The actual derivative evaluations are done in subroutine FXY. Function RKG also includes the F/M control logic. Counter ICNT in function RKG permits transfer to the main program for data output.

TABLE 15. TRANSIENT-STATE ACTIVATED SLUDGE, COMPUTER PROGRAM

```
C     PROGRAM SIMUL
C     TRANSIENT STATE ACTIVATED SLUDGE SIMULATION
C
      DIMENSION Y(2),F(2)
      REAL K,KSUBS,KENDOG
      COMMON V,Q,SFEED,R,XR,K,KSUBS,YIELD,KENDOG,COMPAC
    5 FORMAT(11F)
    6 FORMAT(25X,'F/M CONTROL TEST SIMULATION'//)
    7 FORMAT(22X,'BLANK SIMULATION - NO F/M CONTROL'//)
      LOGICAL RKG,LOGRKG
    8 FORMAT(27X,'KINETIC MODEL CONSTANTS'/14X,'K,1/DAY',
     1'   KSUBS,MG/L TOD    YIELD   KENDOG,1/DAY'/
     215X,F4.1,7X,F6.1,11X,F4.2,5X,F5.3//)
    9 FORMAT(1X,'INPUT VALUES:'/6X,'AERATION BASIN VOLUME',
     1' =',E11.4,' GALLONS'/6X,'INFLUENT FLOW RATE    =',
     2E11.4,' GPM'/6X,'BIO-SETTLER UNDERFLOW =',E11.4,
     3' ML/MIN'/6X,'INITIAL RECYCLE FLOW  =',E11.4,
     4' ML/MIN (RECYCLE RATIO =',F6.3,')'//4X,'INITIAL ',
     5'CONDITIONS:'/6X,'AERATION BACTERIA      =',F8.1,
     6' MG/L MLVSS  (',F7.1,' MG/L TC)'/6X,'SLUDGE BACTERIA',
     7'         =',F8.1,' MG/L MLVSS  (',F7.1,' MG/L TC)'/
     86X,'BIO-SETL OVERFLOW TOD =',F8.1,' MG/L TOD (AER',
     9'ATION SUBSTRATE CONC)'/6X,'INFLUENT WASTE CONC.  =',
     1F8.1,' ML/L TOD (',F6.1,' MG/L TC)'//
     16X,'SLUDGE COMPACTION RATIO =',F5.2,' MG/L SLUDGE PER'
     2,' MG/L BACTERIA'/6X,'TC CONVERSION FACTOR    =',
     3F5.2,' TC/MLVSS'/6X,'TC CONVERSION FACTOR    =',F5.2,
     4' FEED TC/FEED TOD'//6X,'INITIAL F/M LOADING =',
     5F5.3,' TC/TC'///'INITIAL STEADY STATE VALUES AS CAL',
     6'CULATED BY THE KINETIC MODEL:'/1X,'MIXED LIQUOR SUB',
     7'STRATE CONCENTRATION:',F6.1,' MG/L TOD  OR',F7.1,
     8' MG/L TC'/1X,'MIXED LIQUOR BACTERIA CONCENTRATION:',
     9F7.1,' MG/L VSS  OR',F7.1, ' MG/L TC'////6X,'TIME AT',
     1' WHICH LOADING IS CHANGED:',F8.1,' HOURS'/6X,
     2'NEW (T=',F4.1,') INFLUENT WASTE CONC:',F8.1,
     3' MG/L TOD (',F5.0,' MG/L TC)'//)
   10 FORMAT(4G)
   15 FORMAT(2F)
   16 FORMAT(20X,'ACTIVATED SLUDGE TRANSIENT SIMULATION'//)
   17 FORMAT(2F)
   18 FORMAT(4G)
   19 FORMAT('TIME   BACTERIA,MG/L  SUBSTRATE  MIXED ',
     1'LIQ  F/M  RECYCLE  WASTE    WSUM'/'HOURS  M',
     2'LVSS   TC      TOD, MG/L   MG/L TC  TC/TC',
     3'  ML/MIN  ML/MIN  GM VSS')
   20 FORMAT(F5.1,F8.1,F7.1,1X,F8.1,3X,F8.1,
     1F8.3,1X,F7.1,F8.1,1X,F7.1)
```

Table 15 (continued). TRANSIENT-STATE ACTIVATED SLUDGE, COMPUTER PROGRAM

```
C
      READ(1,5)V,Q,SFEED,SFEED2,RECY,W,XR,K,KSUBS,YIELD,KENDOG
      READ(1,10)XSTEP,XF,H,INT
      READ(1,15)Y(1),Y(2)
      READ(1,17)FACTOD,FACVSS
      READ(1,18)LOPT,AA,BB,WMAX
C
      R=RECY/(Q*3875.)
      T=RECY+W
      K=K/24.
      Q=Q*60.
      KENDOG=KENDOG/24.
      FMLOAD=(SFEED*FACTOD)/(Y(2)*FACVSS+Y(1)*FACTOD)
      FEED2C=SFEED2*FACTOD
      COMPAC=XR/Y(2)
      X=0.
      N=2
      WSUM=0.
C
C     CALCULATE INITIAL MODEL STEADY STATE VALUES
      THETA=V/Q
      THETAC=THETA/(1.+R-R*COMPAC)
      SONE=(KSUBS*(1.+KENDOG*THETAC))/
     1(THETAC*(YIELD*K-KENDOG)-1.)
      XSTST=(YIELD*(SFEED-SONE)*THETAC)/
     1((1.+KENDOG*THETAC)*THETA)
      SONETC=SONE*FACTOD
      XSTSTC=XSTST*FACVSS
      TCB=Y(2)*FACVSS
      TCF=SFEED*FACTOD
      XRTC=XR*FACVSS
C
      WRITE(3,16)
      IF(LOPT.EQ.0)GO TO 25
      WRITE(3,6)
      GO TO 26
   25 WRITE(3,7)
   26 K=K*24.
      KENDOG=KENDOG*24.
      WRITE(3,8)K,KSUBS,YIELD,KENDOG
      K=K/24.
      KENDOG=KENDOG/24.
      Q=Q/60.
      WRITE(3,9)V,Q,T,RECY,R,Y(2),TCB,XR,XRTC,Y(1),SFEED,TCF,
     1COMPAC,FACVSS,FACTOD,FMLOAD,SONE,SONETC,XSTST,XSTSTC,
     2XSTEP,XSTEP,SFEED2,FEED2C
      Q=Q*60.
      WRITE(3,19)
```

Table 15 (continued). TRANSIENT-STATE ACTIVATED SLUDGE,
COMPUTER PROGRAM

```
   30 CONTINUE
      LOGRKG=RKG(N,X,XSTEP,SFEED,SFEED2,Y,XF,H,INT,
     1TCB,FACVSS,FACTOD,FMLOAD,LOPT,AA,BB,W,RECY,WSUM,THETA,T,
     2WMAX,R,V,XR,TCMIX)
      WRITE(3,20)X,Y(2),TCB,Y(1),TCMIX,FMLOAD,RECY,W,WSUM
      IF(.NOT.LOGRKG)GO TO 30
      STOP
      END
C
C
      LOGICAL FUNCTION RKG(N,X,XSTEP,SFEED,SFEED2,Y,XF,H,INT,
     1TCB,FACVSS,FACTOD,FMLOAD,LOPT,AA,BB,W,RECY,WSUM,THETA,T,
     2WMAX,R,V,XR,TCMIX)
      REAL K(2)
      DIMENSION Q(2),Y(2),F(2)
      DATA A1,A2,A3,A4,A5,A6/.29289321881345,0.58578643762690,
     10.12132034355964,1.70710678118655,3.41421356237310,
     14.12132034355964/
      DATA ICNT/0/
C     FUNCTION RKG SOLVES A SYSTEM OF N SIMULTANEOUS FIRST
C     ORDER ORDINARY DIFFERENTIAL EQUATIONS USING THE FOURTH
C     ORDER RUNGE-KUTTA METHOD WITH GILL'S COEFFICIENTS.
   40 CONTINUE
      IF(X.GE.XSTEP) SFEED=SFEED2
      CALL FXY(X,Y,F)
      DO 50 I=1,N
      K(I)=F(I)*H
      Y(I)=Y(I)+K(I)*.5
   50 Q(I)=K(I)
C
      X=X+.5*H
      CALL FXY(X,Y,F)
      DO 51 I=1,N
      K(I)=F(I)*H
      Y(I)=Y(I)+A1*(K(I)-Q(I))
   51 Q(I)=A2*K(I)+A3*Q(I)
C
      CALL FXY(X,Y,F)
      DO 52 I=1,N
      K(I)=F(I)*H
      Y(I)=Y(I)+A4*(K(I)-Q(I))
   52 Q(I)=A5*K(I)-A6*Q(I)
C
      X=X+.5*H
      CALL FXY(X,Y,F)
      DO 53 I=1,N
      K(I)=F(I)*H
   53 Y(I)=Y(I)+K(I)/6.-Q(I)/3.
```

Table 15 (continued). TRANSIENT-STATE ACTIVATED SLUDGE, COMPUTER PROGRAM

```
C
C       DO F/M CONTROL OF RECYCLE AND WASTE IF OPTION LOPT=1
        TCB=Y(2)*FACVSS
        TCF=SFEED*FACTOD
        TCMIX=TCB+FACTOD*Y(1)
        FMLOAD=TCF/TCMIX
        IF(LOPT.NE.1)GO TO 56
        W=BB-AA*FMLOAD
        IF(W.GT.WMAX)W=WMAX
        IF(W.LT.0.)W=0.
        RECY=T-W
        R=(0.060*THETA*RECY)/(3.875*V)
     56 WSUM=W*H*XR*.06*.001+WSUM
C
        IF(X.LT.XF) GO TO 60
        RKG=.TRUE.
        RETURN
C
C       PRINT VALUES EVERY INT STEPSIZES.
     60 ICNT=ICNT+1
        IF(ICNT/INT*INT.NE.ICNT) GO TO 40
        RKG=.FALSE.
        RETURN
        END
C
C
        SUBROUTINE FXY(X,Y,F)
        DIMENSION Y(2),F(2)
        REAL K,KSUBS,KENDOG
        COMMON V,Q,SFEED,R,XR,K,KSUBS,YIELD,KENDOG,COMPAC
        XR=COMPAC*Y(2)
        F(1)=(Q/V)*(SFEED-Y(1))-K*Y(1)*Y(2)/(KSUBS+Y(1))
        F(2)=(Q/V)*(R*XR-(1.+R)*Y(2))+YIELD*K*Y(1)*Y(2)
      1/(KSUBS+Y(1))-KENDOG*Y(2)
        RETURN
        END
*
```

Table 16. ACTIVATED SLUDGE TRANSIENT SIMULATION
F/M CONTROL TEST SIMULATION

```
                        KINETIC MODEL CONSTANTS
            K,1/DAY   KSUBS,MG/L TOD     YIELD   KENDOG,1/DAY
              5.0         117.0            .21       .060


 INPUT VALUES:
      AERATION BASIN VOLUME =  .2500E 03 GALLONS
      INFLUENT FLOW RATE    =  .6000E 00 GPM
      BIO-SETTLER UNDERFLOW =  .7850E 03 ML/MIN
      INITIAL RECYCLE FLOW  =  .7000E 03 ML/MIN (RECYCLE RATIO =  .301)

    INITIAL CONDITIONS:
      AERATION BACTERIA     =  1938.0 MG/L MLVSS  (  736.4 MG/L TC)
      SLUDGE BACTERIA       =  7510.0 MG/L MLVSS  ( 2853.8 MG/L TC)
      BIO-SETL OVERFLOW TOD =   116.0 MG/L TOD (AERATION SUBSTRATE CONC)
      INFLUENT WASTE CONC.  =  1556.0 ML/L TOD ( 404.6 MG/L TC)

      SLUDGE COMPACTION RATIO = 3.88 MG/L SLUDGE PER MG/L BACTERIA
      TC CONVERSION FACTOR    =  .38 TC/MLVSS
      TC CONVERSION FACTOR    =  .26 FEED TC/FEED TOD

      INITIAL F/M LOADING = .528 TC/TC


INITIAL STEADY STATE VALUES AS CALCULATED BY THE KINETIC MODEL:
 MIXED LIQUOR SUBSTRATE CONCENTRATION: 116.7 MG/L TOD  OR   30.3 MG/L TC
 MIXED LIQUOR BACTERIA CONCENTRATION: 1992.0 MG/L VSS  OR  757.0 MG/L TC


      TIME AT WHICH LOADING IS CHANGED:     9.0 HOURS
      NEW (T= 9.0) INFLUENT WASTE CONC:  2470.0 MG/L TOD ( 642. MG/L TC)


TIME   BACTERIA,MG/L    SUBSTRATE  MIXED LIQ   F/M   RECYCLE   WASTE    WSUM
HOURS  MLVSS    TC      TOD, MG/L  MG/L TC    TC/TC  ML/MIN   ML/MIN   GM VSS
  3.0  1952.4  741.9      121.0      773.4    .523    709.9     75.1     98.6
  6.0  1962.9  745.9      120.1      777.1    .521    706.7     78.3    203.5
  9.0  1970.2  748.7      119.2      779.7    .519    704.6     80.4    312.4
 12.0  2101.1  798.4      268.7      868.3    .740    785.0       .0    312.7
 15.0  2257.0  857.7      265.7      926.7    .693    785.0       .0    312.7
 18.0  2418.5  919.0      231.3      979.2    .656    785.0       .0    312.7
 21.0  2579.8  980.3      197.3     1031.6    .623    785.0       .0    312.7
 24.0  2739.1 1040.9      170.8     1085.3    .592    785.0       .0    312.7
 27.0  2883.6 1095.7      151.5     1135.1    .566    763.2     21.8    328.9
 30.0  2985.3 1134.4      139.7     1170.7    .549    741.7     43.3    397.4
 33.0  3055.3 1161.0      132.8     1195.5    .537    727.5     57.5    504.8
 36.0  3103.2 1179.2      128.4     1212.6    .530    718.0     67.0    639.3
 38.9  3136.0 1191.7      125.7     1224.4    .525    711.6     73.4    792.5
 39.0  3136.6 1191.9      125.6     1224.6    .524    711.5     73.5    795.7
```

APPENDIX D

STEADY-STATE ACTIVATED SLUDGE COMPUTER MODEL

A computer program was developed to determine the minimum volume of aeration basin required under steady-state conditions for a specified treatment efficiency. Several treatment efficiencies may be specified and the results tabulated according to recycle ratio and recycle sludge concentration. The basis for the calculations is the Monod kinetic model. A good discussion of the development of the model equations is presented by Lawrence and McCartly, 1970.

The input data needed for the execution of this program are:

CARD NO. 1

Q - influent feed flow rate (gpm)
SFEED - influent feed concentration (mg/l TOD)
K - substrate utilization coefficient (day^{-1})
KSUBS - half velocity coefficient (mg/l TOD)
YIELD - growth yield coefficient (1 lb VSS formed/lb TOD removed)
KENDOG - microorganism decay coefficient (day^{-1})

CARD NO. 2

NI - number of efficiencies specified
NJ - number of sludge concentrations specified
NL - number of recycle ratios specified

CARD NO. 3

EFF(I) - efficiency of substrate removal $(S_0-S_1)/S_0$. I = 1, NI

CARD NO. 4

XR(J) - recycle sludge concentration (mg/l MLVSS).
J = 1, NJ

CARD NO. 5

R(L) - volumetric recycle ratio. L = 1, NL

The computer program and a sample printout are given in Tables 17 and 18 respectively.

Table 17. STEADY STATE ACTIVATED SLUDGE COMPUTER PROGRAM

```
C   PROGRAM BACT
C   MONOD MODEL STEADY-STATE CALCULATIONS
C
    DIMENSION EFF(20),XR(20),R(20),RECY(20)
    REAL K,KSUBS,KENDOG
C
  10 FORMAT(6F)
  15 FORMAT(///18X,'STEADY STATE ACTIVATED SLUDGE MODEL'////
   123X,'KINETIC MODEL CONSTANTS'/12X,'K,1/DAY   KSUBS,',
   2'MG/L   YIELD   KENDOG,1/DAY'/12X,F4.1,7X,F6.1,5X,
   3F4.2,7X,F5.3/)
  20 FORMAT(10X,'INFLUENT FLOW, GPM     INFLUENT WASTE',
   1' CONCENTRATION, MG/L TOD'/13X,F10.2,20X,F8.1/)
  25 FORMAT(3I)
  30 FORMAT(20F)
  35 FORMAT(1X//1X,'CASE',I3,'.'/8X,'EFFICIENCY OF',
   1' WASTE REMOVAL =',F5.1,'%   (',F5.1,' MG/L EFFLUENT ',
   2'TOD)'/8X,'RECYCLE SLUDGE CONCENTRATION =',
   3F6.0,' MG/L MLVSS'//5X,'REQUIREMENTS TO MEET THE DESIR',
   4'ED REMOVAL:'/8X,'BIOLOGICAL MASS, XV = ',E10.4,
   5' LBS.'/8X,'F/M LOADING RATIO =',F6.2,' LB TOD/',
   6'LB MLVSS-DAY'/8X,'F/M REMOVAL RATIO =',F6.2,' LB TOD/',
   7'LB MLVSS-DAY'/8X,'SLUDGE AGE REQUIRED =',F5.2,' DA',
   8'YS'/8X,'WASHOUT SLUDGE AGE =',F6.2,' DAYS  (MAX',
   9' W =',E10.4,' GPM)'///)
  36 FORMAT(1X,
   1'RECYCLE    RECYCLE    WASTE SLUDGE    BACTERIA   ',
   2'AERATION   HYDRAULIC'/2X,'RATIO      FLOW,GPM  ',
   3' FLOW, GPM       MLVSS,MG/L  VOL, GAL.  RET*N,',
   4' DAYS')
  40 FORMAT(1X,F5.2,4X,E10.4,3X,E10.4,6X,F6.0,3X,E10.4,
   13X,F6.2)
  45 FORMAT(1X///20X,'********************************'/)
  48 FORMAT(////'CASE',I3,'   RECYCLE SLUDGE CONCENTRATION',
   1' =',F6.0//)
C
    READ(1,10) Q,SFEED,K,KSUBS,YIELD,KENDOG
    READ(1,25) NI,NJ,NL
    READ(1,30) (EFF(I),I=1,NI)
    READ(1,30) (XR(J),J=1,NJ)
    READ(1,30) (R(L),L=1,NL)
C
    WRITE(3,15) K,KSUBS,YIELD,KENDOG
    WRITE(3,20) Q,SFEED
    DO 47 I=1,NL
```

Table 17 (continued). STEADY-STATE ACTIVATED SLUDGE COMPUTER PROGRAM

```
      RECY(I)=R(I)*Q
   47 CONTINUE
      M=0
      L=1
      J=1
      I=1
   50 M=M+1
   51 SAER=SFEED-SFEED*EFF(I)/100.
      THETAC=1./(YIELD*K*SAER/(KSUBS+SAER)-KENDOG)
      AERVOL=(1440.*Q*THETAC*(1.+R(L)))/(1.+(R(L)*
     1XR(J)*(1.+KENDOG*THETAC))/(YIELD*(SFEED-SAER)))
      HRT=AERVOL/(Q*1440.)
      AERX=THETAC*YIELD*(SFEED-SAER)/(HRT*(1.+KENDOG*THETAC))
      W=AERX*AERVOL/(THETAC*XR(J)*1440.)
      IF(L.NE.1) GO TO 56
      IF(J.NE.1) GO TO 140
      FMREM=(K*SAER)/(KSUBS+SAER)
      FMLOAD=FMREM*100./EFF(I)
      AGEMIN=1./((YIELD*K*SFEED/(KSUBS+SFEED))-KENDOG)
      XV=AERVOL*AERX*.001*3.875/454.
      WMAX=AERX*AERVOL/(AGEMIN*XR(J)*1440.)
      WRITE(3,35)M,EFF(I),SAER,XR(J),XV,FMLOAD,FMREM,THETAC,
     1AGEMIN,WMAX
   55 WRITE(3,36)
   56 WRITE(3,40)R(L),RECY(L),W,AERX,AERVOL,HRT
      L=L+1
      IF(L.LE.NL)GO TO 51
      L=1
      J=J+1
      IF(J.LE.NJ)GO TO 50
      J=1
      L=1
      I=I+1
      IF(I.GT.NI)GO TO 150
      WRITE(3,45)
      GO TO 50
      GO TO 47
  140 WRITE(3,48)M,XR(J)
      GO TO 55
  150 STOP
      END
*
```

Table 18. STEADY-STATE ACTIVATED SLUDGE MODEL

```
                    KINETIC MODEL CONSTANTS
        K,1/DAY   KSUBS,MG/L   YIELD   KENDOG,1/DAY
        5.0        117.0        .21        .060

      INFLUENT FLOW, GPM      INFLUENT WASTE CONCENTRATION, MG/L TOD
                .50                         2000.0


ASE  1.
     EFFICIENCY OF WASTE REMOVAL = 80.0%   (400.0 MG/L EFFLUENT TOD)
     RECYCLE SLUDGE CONCENTRATION = 5000. MG/L MLVSS

   REQUIREMENTS TO MEET THE DESIRED REMOVAL:
     BIOLOGICAL MASS, XV =  .2542E 01 LBS.
     F/M LOADING RATIO =  4.84 LB TOD/LB MLVSS-DAY
     F/M REMOVAL RATIO =  3.87 LB TOD/LB MLVSS-DAY
     SLUDGE AGE REQUIRED = 1.33 DAYS
     WASHOUT SLUDGE AGE =  1.07 DAYS  (MAX W = .3855E-01 GPM)


ECYCLE    RECYCLE      WASTE SLUDGE    BACTERIA     AERATION    HYDRAULIC
RATIO    FLOW,GPM      FLOW, GPM     MLVSS,MG/L   VOL, GAL.   RET*N, DAYS
 .10    .5000E-01      .3112E-01        737.      .4038E 03       .56
 .20    .1000E 00      .3112E-01       1093.      .2725E 03       .38
 .30    .1500E 00      .3112E-01       1393.      .2137E 03       .30
 .40    .2000E 00      .3112E-01       1651.      .1804E 03       .25
 .50    .2500E 00      .3112E-01       1874.      .1589E 03       .22


SE  2   RECYCLE SLUDGE CONCENTRATION = 7500.

ECYCLE    RECYCLE      WASTE SLUDGE    BACTERIA     AERATION    HYDRAULIC
RATIO    FLOW,GPM      FLOW, GPM     MLVSS,MG/L   VOL, GAL.   RET*N, DAYS
 .10    .5000E-01      .2075E-01        965.      .3087E 03       .43
 .20    .1000E 00      .2075E-01       1509.      .1973E 03       .27
 .30    .1500E 00      .2075E-01       1970.      .1512E 03       .21
 .40    .2000E 00      .2075E-01       2365.      .1259E 03       .17
 .50    .2500E 00      .2075E-01       2707.      .1100E 03       .15
```

Table 18 (continued). STEADY-STATE ACTIVATED SLUDGE MODEL

```
CASE  3.
       EFFICIENCY OF WASTE REMOVAL = 90.0%   (200.0 MG/L EFFLUENT TOD)
       RECYCLE SLUDGE CONCENTRATION = 5000. MG/L MLVSS

    REQUIREMENTS TO MEET THE DESIRED REMOVAL:
       BIOLOGICAL MASS, XV =   .3507E 01 LBS.
       F/M LOADING RATIO =   3.51 LB TOD/LB MLVSS-DAY
       F/M REMOVAL RATIO =   3.15 LB TOD/LB MLVSS-DAY
       SLUDGE AGE REQUIRED = 1.66 DAYS
       WASHOUT SLUDGE AGE =  1.07 DAYS  (MAX W = .5318E-01 GPM)


 RECYCLE     RECYCLE      WASTE SLUDGE     BACTERIA     AERATION    HYDRAULIC
  RATIO      FLOW,GPM     FLOW, GPM       MLVSS,MG/L   VOL, GAL.   RET*N, DAYS
   .10      .5000E-01      .3438E-01         767.      .5356E 03       .74
   .20      .1000E 00      .3438E-01        1120.      .3669E 03       .51
   .30      .1500E 00      .3438E-01        1418.      .2897E 03       .40
   .40      .2000E 00      .3438E-01        1674.      .2454E 03       .34
   .50      .2500E 00      .3438E-01        1896.      .2167E 03       .30



CASE  4   RECYCLE SLUDGE CONCENTRATION = 7500.


 RECYCLE     RECYCLE      WASTE SLUDGE     BACTERIA     AERATION    HYDRAULIC
  RATIO      FLOW,GPM     FLOW, GPM       MLVSS,MG/L   VOL, GAL.   RET*N, DAY
   .10      .5000E-01      .2292E-01         994.      .4132E 03       .57
   .20      .1000E 00      .2292E-01        1536.      .2674E 03       .37
   .30      .1500E 00      .2292E-01        1995.      .2059E 03       .29
   .40      .2000E 00      .2292E-01        2388.      .1720E 03       .24
   .50      .2500E 00      .2292E-01        2729.      .1505E 03       .21
```